APPLIED RESEARCH REPORT

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COMPUTER SIMULATION:

SUITABILITY OF NOVA SCOTIA COALS AS FEEDSTOCKS FOR BGC/LURGI, DOW, KRW, TEXACO, and SHELL GASIFIERS.

FINAL REPORT

BY

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by

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Mr. David L. Breton	Process Associate Dow Chemical U.S.A. Houston, Texas
Mr. Neil Richter	Texaco Montebello Research Lab Montebello, California.
Mr. F.A.M. Schrijvers	Shell International Petroleum The Hague, Netherlands
Dr. B.H. Thompson	Head of Gasification Services British Gas plc London, England.

We furthermore wish to thank Mr. Michael Mendelson of JSD Simulation Service Company for information on ASPEN during our aborted attempt to establish computer modeling in Nova Scotia. Perhaps this report will spark interest in computer simulation with research funding agencies that are sympathetic to Canadian participation.

SUMMARY

Coal analyses for three Sydney coals (Prince Mine, Hub Seam; Donkin Tunnel, Harbour Seam; Victoria Junction mix, Phalen Seam, Harbour Seam) were supplied to the following gasification process developers:

1. M.W. Kellogg Company (Houston)
2. DOW Chemical USA (Houston)
3. British Gas Co. (London)
4. Texaco (California)
5. Shell International (The Netherlands)

Each process developer utilized in-house proprietary computer software to model the performance of the Sydney coals in a process simulation of their particular system design.

Only the gasifiers were included in the flow sheets for British Gas/Lurgi, Texaco and Shell. The performance of gas turbine units in an IGCC power plant configuration were included in the models of M.W. Kellogg Co. (KRW) and DOW Chemicals. Results show that all three coals of the Cape Breton Development Corporation are suitable as feedstock for the respective gasifier.

IGCC power plant efficiencies as estimated by KRW and DOW reach 43.6% which is considered to be optimistic. Efficiencies decrease with the coal feed in the order: VJ mix > Donkin > Prince. Efficiencies are dependent on the design considerations, especially the mode of desulphurization.

For processes using coal-water fuel (DOW, Texaco), the efficiencies increase with increasing slurry concentration. KRW offers a lower cost desulphurization scheme but at the expense of greater waste disposal and no resultant elemental sulphur by-product.

The IGCC power plant configurations promise an extremely clean gas turbine fuel resulting in very low atmospheric emissions of sulphur dioxide (SO_2) -- much lower than necessitated by regulatory requirements.

The atmospheric emissions of CO and NO_x will be a function of gas turbine design but should meet all regulatory requirements.

Except for the KRW solid wastes, the slag from other gasifiers should be non-leachable and suitable for use in commercial construction material.

Untreated fuel gas will be "sour", i.e. toxic. The elemental sulphur product from gas desulphurization should yield some 40 kg S per ton of coal.

Our review of computer simulation technology applied to coal gasification revealed a multitude of approaches with reliability dependent on the scope of modeling and the limitations of

proprietary data. The methodology is not standardized industry-wide. The models by the process developers, therefore, vary in reliability, and results have not been optimised.

RECOMMENDATIONS

1. Sydney coal suitability for gasification has been predicted but must now be verified by pilot plant testing. Pilot plant tests would permit optimising the gasifier operation conditions.
2. Computer modeling should be extended to sensitivity studies using different design criteria. For example, the KRW system should be modeled with zinc ferrite desulphurization eliminated. The coal input should be varied to reflect the range of chemistries likely to be encountered.
3. Computer simulation for all five processes should be conducted via a common methodology that is not proprietary to any individual process developer.

RÉSUMÉ

Les analyses de trois charbons provenant de Sydney (de la mine Prince, filon central; du tunnel de Dunkin, filon du port; du mélange de la Victoria Junction, filon de Phalen et celui du port) ont été remises à ces compagnies de développement d'un processus de gazéification:

1. la M.W.Kellogg Company (Houston);
2. la Dow Chemical USA (Houston);
3. la British Gas Co. (Londres);
4. Texaco (Californie);
5. Shell International (Pays-Bas).

Chacune de ces compagnies a utilisé son propre logiciel déposé pour modéliser le rendement des charbons de Sydney, au moyen d'une simulation de fonctionnement fondée sur la configuration particulière de son système.

La British Gas/Lurgi, Texaco et Shell n'ont pris en considération que les gazéificateurs dans le schéma des variables. Les modèles de la M.W. Kellogg Co.(KRW) et de la DOW Chemicals ont tenu compte du rendement des unités de turbines à gaz dans la configuration d'une centrale d'énergie IGCC. Les résultats indiquent que les trois types de charbon de la Société de développement du Cap-Breton peuvent servir de charge d'alimentation pour leur gazéificateur respectif.

Le taux de rendement de la centrale énergétique IGCC, selon l'estimation de la KRW et de la Dow, atteint 43,6%, chiffre considéré optimiste. Le rendement de l'approvisionnement en charbon décroît selon l'ordre suivant: mélange de la VJ > Donkin > Prince. Il dépend

des considérations de design, particulièrement en rapport avec le mode de désulfuration.

Dans les procédés qui utilisent le combustible eau-charbon (DOW, Texaco), le rendement augmente, de même que la concentration de bouillie. La KRW offre un processus de désulfuration à moindre coût mais qui produit davantage de déchets, sans sous-produit de soufre élémentaire.

Les configurations de la centrale énergétique IGCC laissent entrevoir la possibilité d'un combustible de turbine à gaz extrêmement propre, causant des émissions atmosphériques de bioxyde de soufre (SO_2) très faibles, bien en deçà des limites réglementaires.

Les émissions atmosphériques de CO et de NO_x dépendront du design des turbines à gaz mais elles devraient se conformer aux limites prescrites.

À l'exception des déchets solides de la KRW, les scories provenant d'autres gazéificateurs ne devraient pas produire de fuites et devraient pouvoir entrer dans la composition de matériaux de construction commerciaux.

Le gaz combustible non traité sera toxique ("aigre"). La désulfuration du gaz produira du soufre élémentaire au taux approximatif de 40 kg S la tonne de charbon.

Notre examen de la technologie de simulation informatisée telle qu'appliquée à la gazéification du charbon a révélé une foule d'approches, dont la fiablilité est fonction de l'envergure de la modélisation et des limites des données de propriétaire. Cette méthodologie n'est pas normalisée partout dans l'industrie. Par conséquent, les modèles des concepteurs de procédés sont de fiabilité variable et les résultats n'ont pas été maximisés.

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COMPUTER SIMULATION:
SUITABILITY OF NOVA SCOTIA COALS AS FEEDSTOCKS FOR SHELL, TEXACO, BGC/LURGI AND KRW GASIFIERS.

PROJECT HISTORY

Computer process modeling of gasification for Nova Scotia coals was first suggested by the Atlantic Coal Institute during a major review of gasification technology funded by the Nova Scotia Department of Mines and Energy, Mines and Resources Canada (June 1987 progress report, Gasification Project Technical Committee, Canada-Nova Scotia Oil Substitution and Conservation Agreement). A proposal for computer modeling via the commercial package ASPEN (Advanced Systems Process Engineering) was submitted to the Committee on Atlantic Coal on February 12, 1988 by Geofuel Research Inc. The company planned to establish the process modeling capability on microcomputers located in Sydney, Nova Scotia. The Committee on Atlantic Coal identified this project as a priority for potential funding (letter, March 21, 1988).

A less ambitious proposal was submitted to the Cape Breton Development Corporation (CBDC) by the Atlantic Coal Institute on October 20, 1988 which formed the present contract issued by CBDC on March 16, 1989 (Purchase Requisition #15058-ATC10). This contract shifted the emphasis from in-house computer modeling in Sydney to process modeling conducted via subcontracts to the international gasification pilot plant developers. The Appendix contains data generated via proprietary software and procedures

unique to each developer. The level of interpretation in this report is controlled by the level of access to proprietary information.

IGCC TECHNOLOGY

The Atlantic Coal Institute completed a major review of gasification and IGCC technology up to November 1988, as pertains to Nova Scotia coal in a report entitled:

> " Coal Gasification: Manufacture of a Clean Gas for Electric Power Generation in the Province of Nova Scotia"
> (Anderson et al. 1989).

The present report assumes that the reader has access to this report or a general understanding of IGCC technology. Other reviews of IGCC technology can be found in the papers by Gluckman and Louks (1982) and Stochl and Nainiger (1984). A generalized block flow diagram for IGCC in given in Figure 1.

MATHEMATICAL MODELING

Buekens and Schoeters (1983) have reviewed mathematical modeling as pertains to gasification technology. In the process of gasification, there are physical phenomena (gas and solids flow, heat and mass transfer) and chemical phenomena (thermal decomposition, combustion, gasification): these occur simultaneously with numerous interactions. In each type of gasifier, there is a sequence of phenomena (drying, heating, decomposition, gasification, combustion) with rates and sequence dependent on the type of reactor and feedstock.

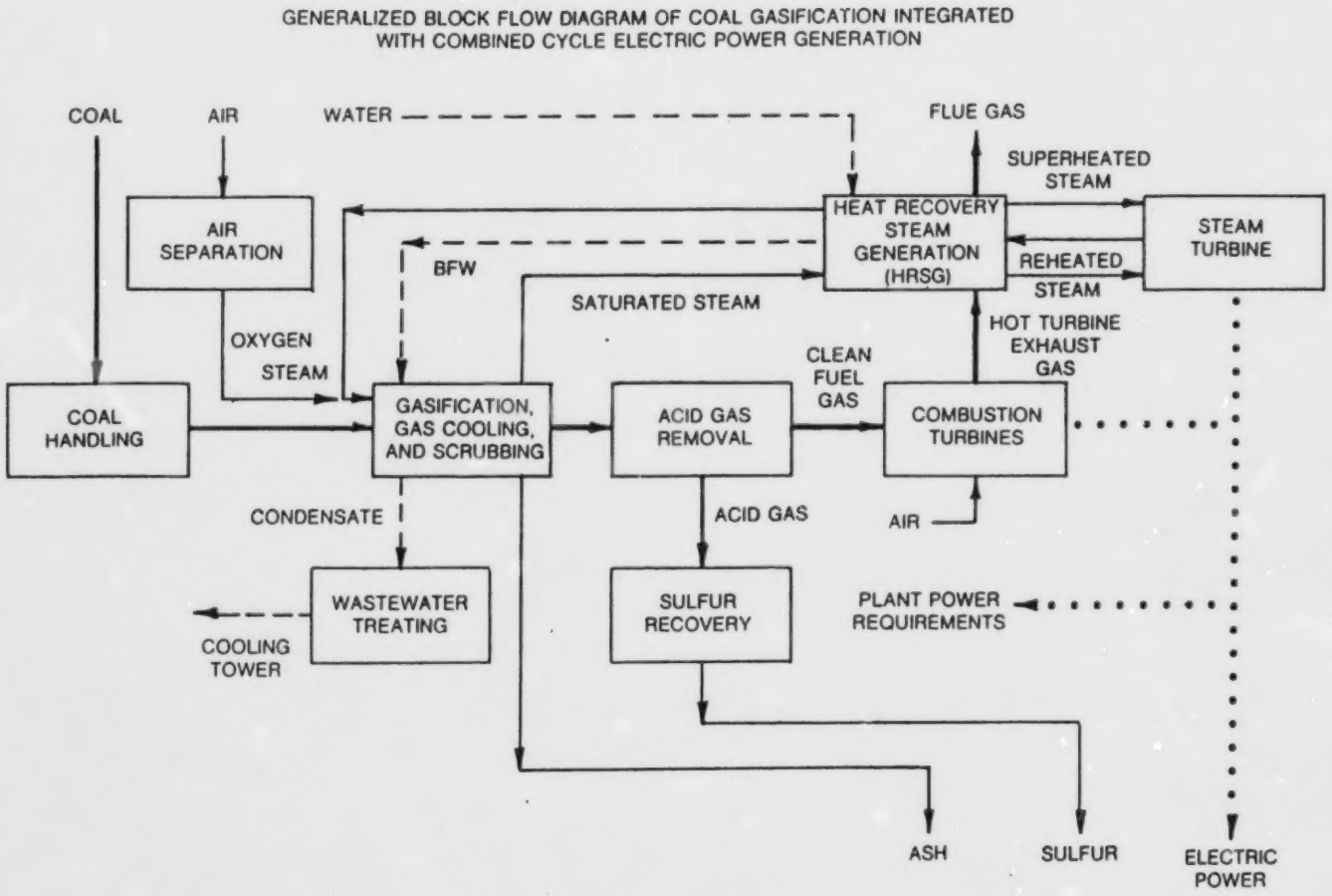


FIGURE 1. GENERALIZED BLOCK FLOW DIAGRAM OF COAL GASIFICATION INTEGRATED WITH COMBINED-CYCLE ELECTRIC POWER GENERATION

(From: Stefano (1985))

The fuel properties having the greatest effect on gasification include:

proximate analysis
ultimate analysis
particle size
thermal behaviour of fuel and residue
higher and lower heating values

Low ash content reduces sintering and slagging, particle size controls the heating rate and mass transfer, while elemental analysis (C,H,N,O,S) indicates the potential for corrosive or polluting gases.

Operating parameters affect the temperature, pressure and residence time. A rise in temperature enhances the rate of all chemical and physical reactions. The temperature, however is not an independent variable: it depends on fuel quality, partial combustion and reactor heat losses. Pressure influences the rates of reaction, heat and mass transfer and modifies the equilibrium relationships. Residence time is a major parameter in gas phase reactions such as thermal cracking. For gas/solids reactions the residence time may refer to "average residence", or that of a bubble or gas phase. The residence time of solids is generally controlled by the flow rate of the blast.

The air factor is important for fluidized bed gasification. High air factors yield gas with low heating value, low air factors give low reactor temperatures and low gasification rates.

Mathematical models may be empirical (if little data is available) or mechanistic. Mechanistic models are composed of conservation equations for mass, energy and momentum. A mass balance is written for each component entering or leaving the

reactor: e.g. C, O_2, H_2O, CO, CO_2, H_2, CH_4, H_2S. The rate of consumption or formation is given by rate equations derived by kinetic experiments.

In simple models it is assumed that specific reactions attain thermodynamic equilibrium. Many models neglect the difference in temperature between gas phase and reacting particle. The momentum balance is generally neglected which implies that the reactor operates under isobaric conditions. This approximation disregards the pressure drop occurring over a fluidized-bed or over a moving bed operating on fine material. Single particle models using the shrinking core concept for combustion may take into account that oxygen diffusion from the particle surface is the rate controlling step whereas simplified models assume combustion is instantaneous. At lower temperatures in a fluidized bed gasifier the chemical reactions may not be limited to particle surfaces. Also, the flow imperfections occurring in a moving bed gasifier such as channelling, bridging and peripheral fire, have never been modeled (Buekens and Schoeters, 1983).

In kinetics-free models, the reactor is subdivided into a number of zones, each featuring a dominant phenomenon such as: drying, pyrolysis, gasification, combustion. The corresponding temperature and concentration gradients are more easily simulated by computer than verified experimentally. Therefore the relative merits of various models are difficult to check in the absence of control data.

Stochl and Nainiger (1984) conducted a comparison of three of the generic types of gasifiers included in our study: British Gas/Lurgi slagging gasifier; Texaco entrained-bed; KRW fluidized-bed gasifier. Their simplified model equations are given in the Appendix. Govind and Shah (1984) and Schneyer et al. (1982) present detailed equations for an entrained flow coal gasifier (Texaco-type). Mori et al. (1984) published equations for a fluidized-bed coal gasifier developed in Japan. It is beyond the scope of our study to participate in these mathematical gymnastics. Furthermore we have no information on the specific mathematical models used by our subcontractors.

Four categories of computer models are available to characterize a KRW type fluidized-bed gasifier (Blake and Liss, 1985): stoichiometric; classical; hydrodynamic; dimensional.
The stoichiometric models are used to predict material and energy balances and process efficiency. Classical models predict the influence of hydrodynamic mixing on the mass balances and efficiency while hydrodynamic models predict the mixing. Dimensional analysis correlates and scales experimental data.
Blake and Liss (1985) conclude that:

> *"...Future development of any of the models is of limited value, unless there is also the development of a database against which the models can be compared. At the present time, the data base is rather limited; there is data for hydrodynamic mixing in a nonreactive environment, for material and energy balances of PDU and laboratory scale reactors, and for quantitative observations of ash agglomeration and agglomerates removal. In the absence of an*

improvement in this database, future model development should be limited to those aspects of modeling, such as improved algorithms in the hydrodynamic models or process specific chemistry in the stoichiometric models which do not depend on new data for model verification."

MODELING SOFTWARE

INTRODUCTION

Process modeling by computer is common to many areas of petroleum science and chemical engineering with a plethoria of software available for general application or specific problem solving. Most programmes are not suitable for investigating coal-based processes because they do not handle solids. Therefore specialized packages have been written for coal conversion processes of which most are proprietary but some are commercially available.

The Electric Power Research Institute (1987) published a guide to computer programs that list the following software with potential application to IGCC technology:

1. UNIRAM: Power Generation System Availability Assessment Model: used to assess IGCC power plants (EPRI AP-2202, 1982).

2. COGEN3: Design Costing and Economic Optimization of Cogeneration Projects. Mathtech Inc.

Other commercial programmes that may be used for modeling some or all of an IGCC power plant include:

3. MASSBAL MK2 and CADSIM: general simulation software by SACDA Inc. (University of Western Ontario).

4. HYSIM: Modeling for gas systems by Hyprotech Ltd. (Calgary, Alberta).

5. BILCO: Material balance modeling by the Bureau de Recherches Geologiques et Minieres, France.

6. CHEMCAD, TOWER+ : Process simulation and hydrocarbon distillation simulation by Coade Engineering Software , Houston, Texas.

7. ASPEN: Process simulation for coal conversion technology by the Massachussetts Institute of Technology.

8. ARACHNE Process Flowsheeting Package by the British Coal Corporation.

The published literature makes references to numerous other mathematical models which may be accessible as software or may be proprietary to the authors (e.g.):

1. S-CUBE/EPRI: EEF MODEL for Texaco entrained bed gasifiers.

2. Govind and Shah (1984): model for the Texaco entrained-bed gasifier.

3. MAINADAMS.DAT: (Cheng and Talbot, 1986) model for entrained-bed gasifiers for phase-plane analysis.

4. Modified Bubble Assemblage Model (Mori et al. 1984): model for fluidized-bed coal gasifier.

It is beyond the scope of the present study to review all of these software options. Because of the proprietary nature of the computer modeling conducted for our study by the process developers, it was not possible to assertain the nature of the software or algorithms used. Therefore our discussion will summarize only the available published information. We assume that some of the output data received for this study were derived with ASPEN, S-CUBE/EPRI (EEF), or ARACHNE software; therefore, these will be

discussed in more detail. On the otherhand, all the process developers claim to use in-house proprietary programs.

ADVANCED SYSTEMS PROCESS ENGINEERING (ASPEN)

ASPEN has been used extensively in coal conversion simulations (Stone, 1986, Becker et al. 1986, Rinard, et al. 1986; Stefano, 1985). It was developed at the Massachusetts Institute of Technology since 1976 under contract to the U.S. Department of Energy (DOE) to simulate coal processes and is now sold as a commercial software package. Britt et al. (1981) cite more than 250,000 lines of software code.

The first step in simulating a process is to develop an ASPEN model of the flowsheet in terms of blocks and streams. ASPEN first solves the heat and material balance relationships (non-linear algebraic equations) to determine values of intermediate and product stream variables. Since the process may involve recycle streams and design specifications, the heat and material balance requires convergence of an iterative procedure. Once the flowsheet is converged, the system computes performance variables for each process unit, the sizes of equipment and operating costs of the plant.

Several gasification models have been developed with ASPEN to date (Stefano, 1985):

1. RGAS - a dry-ash fixed-bed gasifier model developed by Halcon/Scientific Design Company specifically for lignite. Two kinetic routines are available.

2. USRWEN - dry-ash fixed-bed gasifier model developed by C.Y. Wen at West Virginia University. Kinetic model available for bituminous, subbituminous and lignite.

3. CONOCO/MIT - slagging gasifier model developed by Massachusetts Institute of Technology based on Continental Oil Company's design study for a BGC/Lurgi type slagging gasifier. Non-kinetic model with limestone flux, using the Redlich-Kwong-Soave equation of state.

4. ORNL dry-ash gasifier model developed by Oak Ridge National Laboratory. Nonkinetic model.

5. SLAGGER - modified MIT model developed by Stefano (1985) for fixed-bed gasifiers.

ASPEN contains a databank for the physical properties of many compounds and includes equations of state, mass balance, economics etc. A block diagram is developed to characterize a process and input the chemical operation models. Not every physical operation that exists in a process is included in the simulator. Therefore more than one simulation block is usually developed to model a processing step. For this reason the block diagram approximates the process operation but does not necessarily match the process flowsheet.

Therefore computer process modeling, even with this software package is a major undertaking. For example, the simulation of the CFB Summerside AFBC project required an ASPEN input file of 924 lines as follows (Becker, et al. 1986): 39 unit operation models; 4 user developed Fortran programs 8 design specifications. For the Great Plains Gasification project, Rinard et al. (1986) developed 11 ASPEN models simulating fourteen sections of the plant.

Accurate thermodynamic data for each chemical species is necessary and the equations of state used for the model must be known. For gases the following thermodynamic constants are necessary: molar mass; freezing point; boiling point; critical temperature, pressure, volume; critical compressibility; polynomial heat capacity constants; standard heat of formation; Gibbs free energy of formation. For solids the parameters needed are: molar mass; standard heat of formation; Gibbs free energy of formation; polynomial heat capacity constants; polynomial molar volume constants. For our study the proprietary nature of the models did not allow us to verify the thermodynamic data.

All models have inherent weakness due to simplifications in the form of assumptions. The USRWEN gasifier model, for example (Stefano, 1985) makes the following approximations:

1. Devolatilization is assumed to occur separately from other reactions at the top of the gasifier.
2. All gas-solid reactions are represented by volumetric reactions with combustion occuring according to the shrinking core model.
3. There is no difference between gas and solid temperatures at any point in the gasifier.
4. The temperature at the bottom of the gasifier is assumed equal to inlet temperature of the oxidant and steam.
5. The amount of volatile matter produced is specified by the user and doesn't change with operating conditions.
6. The C/H/O ratio in tar is same as in feed coal.
7. Nitrogen forms only N_2 gas, sulphur forms only H_2S gas.
8. Devolatilization is assumed to be thermally neutral.

Rinard et al. (1984) identified shortcomings in ASPEN when applied to simulation of the Great Plains coal gasification plant. Specifically they mentioned the absence of intermediate level models for separation operations located downstream from the gasifier. Furthermore they noted that the data bank was "for the most part reliable" but lacked documentation.

ARACHNE PROCESS FLOWSHEETING PACKAGE

ARACHNE (A Reactor And Cycle Handling Network) was developed in APL language by the National Coal Board of Britain since 1980 to handle coal conversion processes. Emphasis has been on mass and energy balances for coal gasification for the production of hydrogen and other synthetic fuels. The programme is suitable for coal gasification or combustion but cannot model direct liquefaction routes. A costing model (IDMON) is included to allow economic evaluation of process options assessed technically by ARACHNE. The total package is available commercially for 100,000 pounds.

Coal gasification processes that have been addressed include entrained phase, fixed bed and fluidized-bed gasifiers. Models have been developed for two second-generation gasifiers:

1. Texaco entrained-bed
2. British Gas/ Lurgi slagging fixed-bed gasifier.

ARACHNE, as with other flowsheeting programmes, contains a library of "modules" representing operations common to many processes such as: compressors, turbines, heat exchangers, stream

splitters, separators and mixers (Holmes et al., 1981). Additional modules can be created using processor programs such as REACTOR (model chemical reactors), CALCULATOR (model stream data and calculate parameters). Components are then assembled in a "network" with the additional input of physical properties data.

Homes et al. (1981) applied ARACHNE to combined-cycle power generation for production of low-BTU gas.

EPRI ENTRAINED FLOW (EEF) MODEL (S-CUBE/EPRI)

Schneyer et al. (1982) published theoretical formulations along with calculations for three different approximations of the Texaco reactor design. This model was applied to gasification-combined-cycle power generation at the Texaco Cool Water Demonstration Project. The Shell coal gasification process also uses single-stage, pressurized, slagging entrained flow reactors. While the Texaco process requires a coal-water slurry feed, the Shell process uses a dry coal feed. The EEF model with modifications could accomodate the Shell process.

The EEF model is capable of simulating a wide variety of asymmetric geometrical configurations for cylindrical reactors. Calculations include gas-solid-liquid phases including the coal-water slurry and oxygen injection used at Texaco.

LIMITATIONS TO MODELING

PROPRIETARY DATA

Due to the proprietary nature of gasification technology only a limited amount of information is in the public domain. This includes schematic flowsheets, brief process descriptions and appropriate material balances. It is therefore not possible to develop a complete description of each section that is modeled (Rinard et al. 1986). Many process parameters such as equipment performance specifications, recycle flow rates, operating temperatures and pressures are not available. This inhibits the comparison of outputs from the various process developers with published computer models.

Schneyer et al. (1982) pointed out the limitations caused by proprietary data with respect to modeling the Texaco entrained flow gasifier:

> *"... The lack of access on this project to proprietary data has limited the model validation capability... simulations imply that the Texaco class of reactors is rather insensitive to many variations in geometrical design and flowrates and relatively straightforward in operation... The calculations, however reveal a hint of a more complex pattern of operation, development of which would be dependent on coal volatile content, slurry composition, and oxygen flowrate. Such a development in the real reactor might modify or even negate several of the tentative conclusions. ...Validation of the overall model must involve use of Texaco-proprietary information."*

SOFTWARE BUGS

The complexity of computer programs used for process modeling and the addition of user customized routines suggests that software bugs are a significant source of errors. Rinard et al. (1986) discovered a number of "bugs" in the public version of the ASPEN process simulator.

SCOPE OF EFFORT

The level of effort in the design of computer models and the verification of input data and output controls the value of results. Unfortunately the scale of effort of this study is several orders of magnitude below international standards because of limited financing.

The ASPEN model for the CFB Summerside project took two years to develop at a cost of $85,000 (Canadian) (Becker et al. 1986). This represented a highly subsidized effort via university faculty plus access to the software development and experience of the U.S.A. based ASPEN.

Systems Science and Software Inc. (California) received $1,352,776 (U.S.) in 1975 to develop a computer model for fluidized-bed gasification. The British Coal Corporation estimated a time of 300 manhours required to define input for design assumptions, flowsheet models etc. on three gasifier types using existing ARACHNE models.

For our project, the level of effort by the process developers is partially subsidized by their commercial interest in promoting their gasifier. Nevertheless, with subcontracts ranging from "free" for SHELL to only $13,400 (Canadian) for British Gas/Lurgi we cannot expect serious in-depth studies that take into account all the factors.

MASS BALANCE

Stefano (1985) presented results from the USRWEN dry-ash gasifier model for Illinois #6 (Table 1.) which is a high-sulphur bituminous coal similar to Sydney coals. The major discrepancies noted were a low prediction of methane and an over estimation of heat loss by an order of magnitude. A similar simulation using the CONOCO/MIT model for Illinois #6 coal yielded erroneous outlet gas composition for N_2 and H_2S. The modified model (SLAGGER) gives more accurate gas composition and allows more flexibility in recycling of heavy hydrocarbons and resembles more closely a slagging gasifier in an IGCC design (Stefano, 1985).

Stone (1986) modeled a BGC/Lurgi dry-ash slagging fixed-bed gasification IGCC power plant of 509 MW using a modified SLAGGER model (Figure 2). This ASPEN simulation tracked 27 chemical components. Good agreement was achieved between reported and calculated fuel gas compositions (Figure 3). The gas turbine fuel gas stream was designed to yield dry-basis HHV of 364 Btu/scf compared to a modeled value of 380 Btu/scf.

Blake and Liss (1985) reviewed process modeling for fluidized-bed gasifiers and concluded that ash agglomeration is not adequately handled by any available model.

COST AND PERFORMANCE

Stefano's (1985) study yielded a power plant thermal efficiency of 39.3% compared to 39.4% reported for the conceptual design. On the otherhand, Gluckman and Louks (1982) compared initial performance estimates for Texaco-based IGCC plants (Figure 4) and reported a range of estimates from 15% to 35%. When a single base-case configuration was employed for burning Illinois #6 coal, thermal efficiencies for four designs ranged from 37.1% to 37.9%. Therefore we expect that thermal efficiency predictions for IGCC power plants are speculative and vary considerably, even for the same system if component and operating parameters are modified. Results from computer modeling must therefore use the same base case configuration for valid comparision.

Models of coal gasification combined cycle plants may fail to properly represent actual power plants in the following:

1. Plant capacity (maximum output in megawatts)
2. Heat rate with ambient temperature

IGCC plants have a more complicated heat rate curve as a function of plant output than conventional coal-steam plants (Zaininger, 1984). Planning models generally use constant values for capacity and heat-rate for thermal units but in reality,

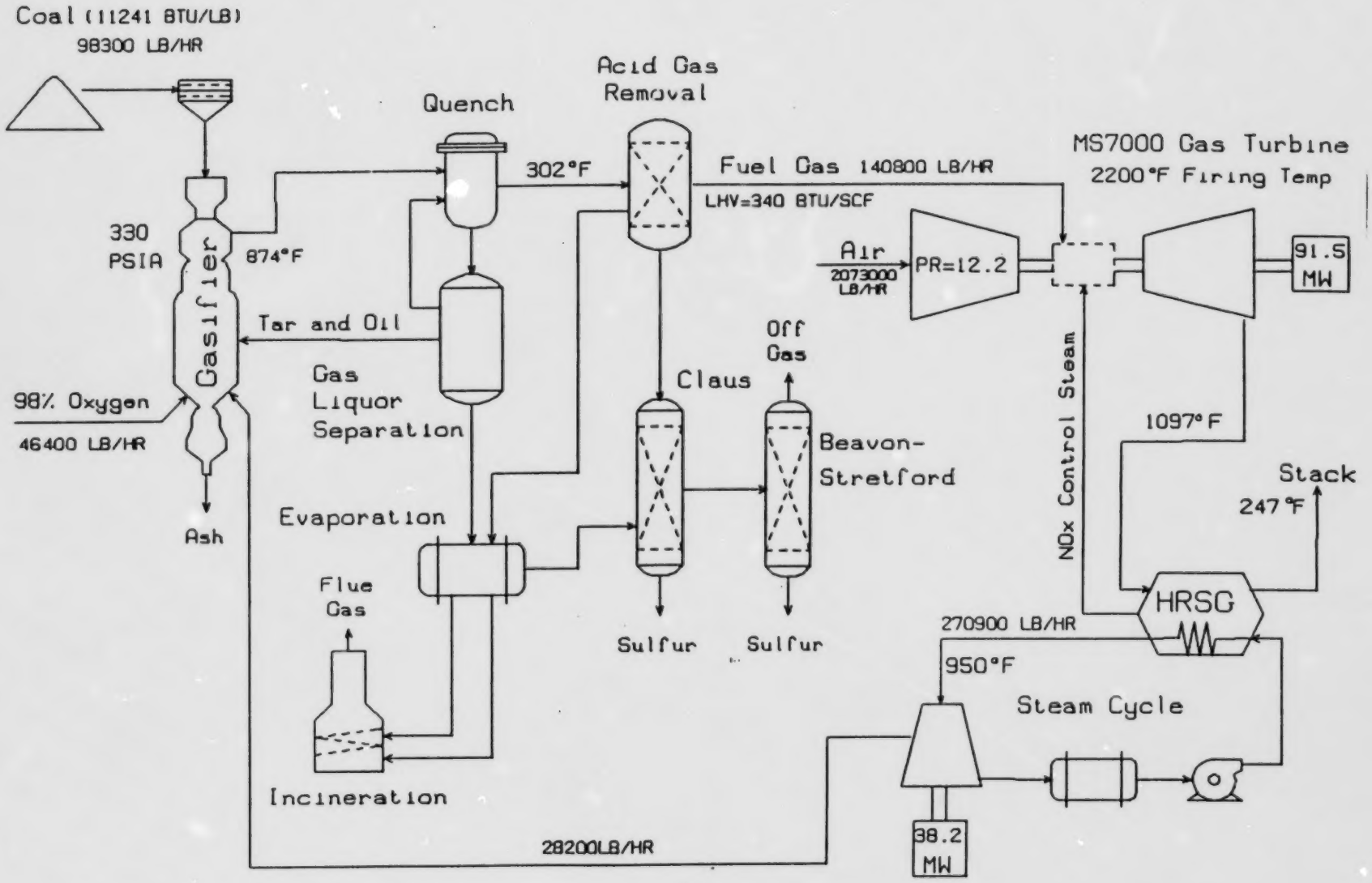


FIGURE 2. BGC/LURGI SLAGGING FIXED-BED GASIFICATION IGCC PLANT
(After Stone (1986) Figure 5)

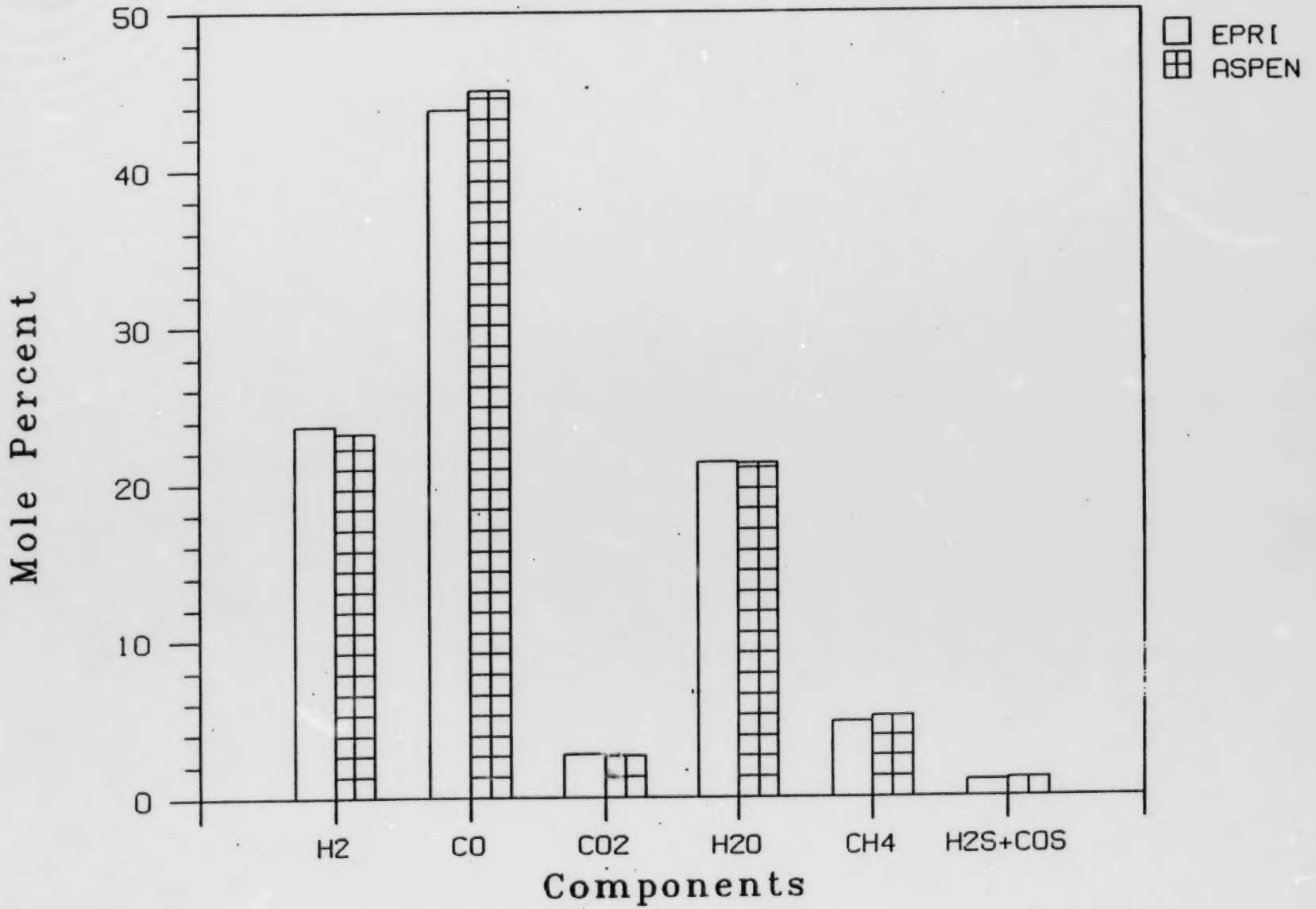
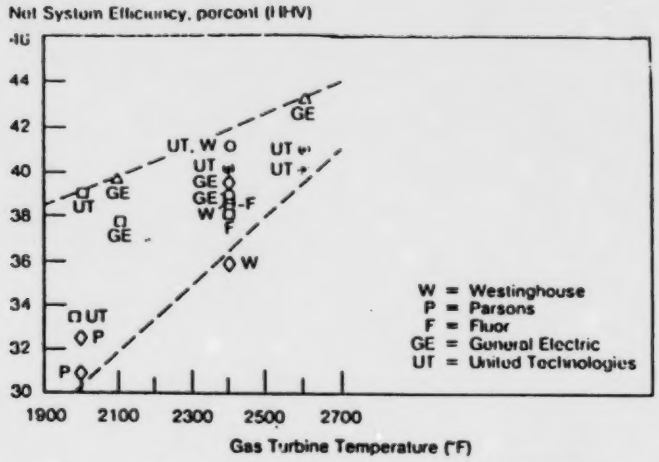


FIGURE 3. RAW GAS COMPOSITION AFTER QUENCH BGC/LURGI SLAGGING FIXED-BED GASIFIER

(From: Stone (1986))



Initial performance estimates for Texaco-based GCC plants.

Figure 4a. After Gluckman and Louks (1982)

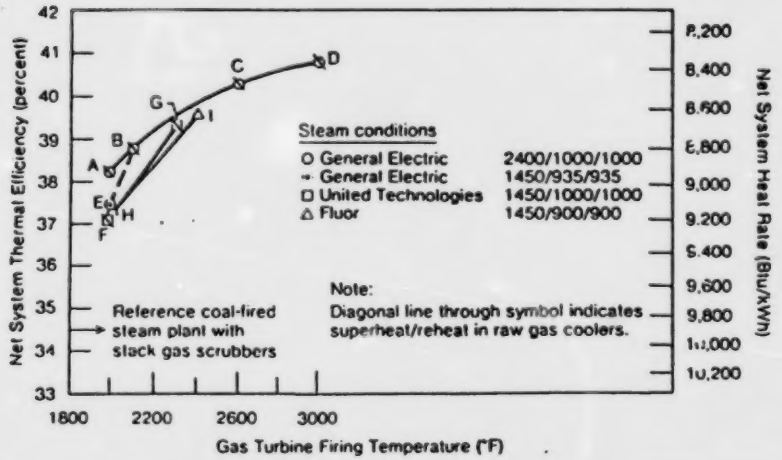


Fig. Effect of gas turbine firing temperature on Texaco-based GCC system performance.

Fibure 4b. After Gluckman and Louks (1982)

these values are a function of ambient temperature. Temperature modeling can be improved by the following refinements:

1. Use site-specific average ambient temperature.
2. Vary performance values throughout the year based on site-specific seasonal changes in temperature.

Zaininger (1984) found significant increases in IGCC energy production when seasonal temperature fluctuations were incorporated. A second difficulty in thermal modeling relates to the unusual heat rate curve: gas turbines exhibit a saw-tooth pattern as they come on-line.

Becker (1986) compared actual AFBC plant operation with ASPEN predictions and concluded that:

1. Heat and material balances match closely.
2. Solids are modelled well.
3. Equipment sizes were smaller or equivalent to those selected by the plant.
4. Pollutant production was predicted well.

SENSITIVITY STUDIES

Computer models can be used to conduct a sensitivity analysis: a study of the operational behaviour of a plant if certain variables are manipulated. Such "what if" modeling leads to economic optimization of the plant design and operations. In the absence of in-house modeling our study was limited to the single-run results supplied by the process developers. No such variation of parameters was therefore possible.

NOVA SCOTIA COALS

EASTERN CANADA PROJECTS

Our study represents the first comprehensive application of process modeling to evaluate Nova Scotia coals as gasification feedstock in IGCC technology. A confidential study by the Scotia Coal Synfuels Project in 1981 included computer simulation of gasifier performance, using Harbour seam coal from the Lingan Mine. Their study, conducted by Texaco Development Corporation (White Plains, New York) focused on methanol production.

Computer simulation has been previously applied to Atmospheric Fluidized Bed Combustion (AFBC) based on an industrial-scale (AFBC) boiler at Canadian Forces Base, Summerside, Prince Edward Island. Dr. H.A. Becker and others at Queen's University applied ASPEN to develop process flowsheets, operating conditions, heat and material balance predictions, and economic analysis (Becker et al. 1986).

COAL CHARACTERISTICS AND SELECTION

Anderson et al. (1989) discussed in detail the coals of the Sydney Basin that would most likely be feedstock for a local IGCC power plant. The selection is based on high-sulphur coals presently being mined (Prince Mine - Hub seam; Phalen Mine-Phalen seam) or forming large blocks of proven reserves (Donkin-Morien Development - Harbour seam). For the purpose of the computer simulation, we assumed that Phalen seam coal would be

mixed with Harbour seam coal at the Victoria Junction wash plant. Therefore a "VJ Thermal" average analysis was utilized.

Appendix 2 identify the coal characteristics requested by each process developer for the purpose of computer modeling. In Appendix 3 we list the data supplied to the modelers for the three coals under consideration. The average coal analyses used for the model are somewhat arbitrary because of the changing nature of coal seams laterally and vertically and the influence of mining methods.

Analyses used for the models do not match the averages reported by Anderson et al. (1989, page 7.4) nor do they reflect the ranges encountered. Instead, they represent "official" values supplied by CBDC (Donkin-Morien, VJ Thermal) and the Nova Scotia Department of Mines (Prince-Hub) based on bulk analyses from other coal utilization studies to allow comparative interpretation. The NSDME data is based on the analyses provided by the Nova Scotia Power Corporation to the Canadian Electrical Association for a Point Aconi IGCC site study by Bechtel of San Francisco. A more rigorous review of gasification models in the future should consider the anticipated range in coal feed composition.

REVIEW OF MODELS

COAL STANDARD: Illinois #6

All the gasification processes have experience with bituminous coals but only Illinous #6 coal has published data for

TABLE 1. ASPEN (USRWEN) results for dry-ash fixed-bed gasifier model: using Illinois No.6 coal. (From: Stefano, 1985)).

Coal Proximate Analysis	Input Values	
Volatile Matter (%, MF basis)	33.96	
Fixed Carbon (%, MF basis)	43.24	
Moisture (%, as received)	13.68	
Ash (%, MF basis)	9.12	
Gasifier Conditions		
Pressure (psia)	365	
Steam/Coal (lb/lb)	2.63	
Oxygen/Coal (lb/lb)	0.50	
Raw Gas Composition (vol %)	**Reference**	**USRWEN**
CO	16.3	17.3
H_2	39.8	46.6
CH_4	8.7	4.8
CxHy	0.6	N/C
H_2S	1.1	0.9
CO_2	32.1	28.9
N_2	1.4	1.6
Gas Heating Value (Btu/scf)	279	270
Offgas Temperature (°F)	1,158	1,122
Tars/Coal (lb/lb)	0.02	0.07
Heat Loss (Btu/hr)	2.62×10^6	5.01×10^7
Cold Gas Efficiency (%)	75.2	79.5
Carbon Conversion (%)	99.7	99.2
CPU Time for Simulation (sec)	N/A	411

N/C -- Not calculated by model.

N/A -- Not applicable.

TABLE 2. SYSTEM SUMMARY OF THE BGC/LURGI SLAGGING FIXED-BED GASIFICATION IGCC PLANT

(From Stone (1986))

*** GASIFIER CONDITIONS ***

COAL FLOW RATE:	0.982991E+05 LB/HR
OXIDANT FLOW RATE:	0.464401E+05 LB/HR
STEAM FLOW RATE:	0.282035E+05 LB/HR
GASIFIER PRESSURE:	330.0 PSIA
EXIT GAS TEMPERATURE:	873.9 F

*** MS7000 GAS TURBINE CONDITIONS ***

FUEL FLOW RATE:	0.140840E+06 LB/HR
AIR FLOW RATE:	0.207297E+07 LB/HR
FUEL LHV (WET):	6552.4 BTU/LB, 340.4 BTU/SCF
FUEL HHV (WET):	6972.8 BTU/LB, 362.2 BTU/SCF
FIRING TEMPERATURE:	2200.0 F
COMBUSTOR EXIT TEMPERATURE:	2306.6 F
TURBINE EXHAUST TEMPERATURE:	1096.5 F
THERMAL EFFICIENCY (LHV):	0.3382
GENERATOR EFFICIENCY:	0.9850

*** STEAM TURBINE CONDITIONS ***

STEAM DIVERTED TO GASIFIER:	0.282035E+05 LB/HR
STEAM TURBINE FLOW RATE:	0.270932E+06 LB/HR
SUPERHEATED STEAM TEMPERATURE:	950.0 F
REHEAT STEAM TEMPERATURE:	900.0 F
EXPANDED STEAM QUALITY:	0.9196
GENERATOR EFFICIENCY:	0.9850

*** POWER PRODUCTION SUMMARY ***

GAS TURBINE:	0.914604E+08	WATTS
STEAM TURBINE:	0.467534E+08	WATTS
COMPRESSORS:	-0.273190E+06	WATTS
PUMPS:	-0.188328E+07	WATTS
OXYGEN PLANT:	-0.883352E+07	WATTS
PLANT TOTAL:	0.127224E+09	WATTS

****
* PLANT THERMAL EFFICIENCY (HHV) = 0.3928 *
****

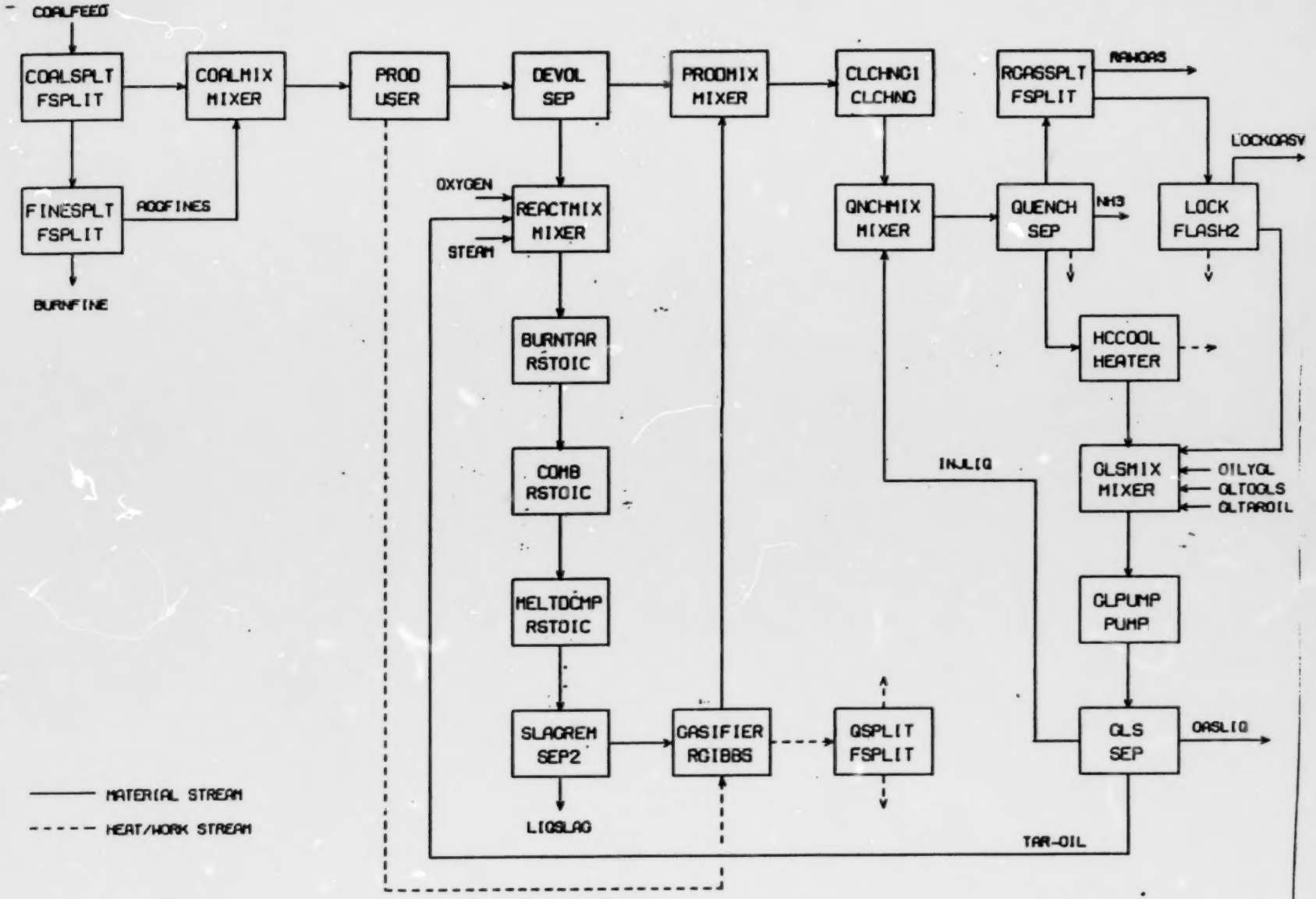


FIGURE 5a **SLAGGER AND LIQSEP FLOWSHEET SECTIONS FROM THE BGC/LURGI SLAGGING FIXED-BED GASIFICATION IGCC PLANT**
(From: Stone, (1986))

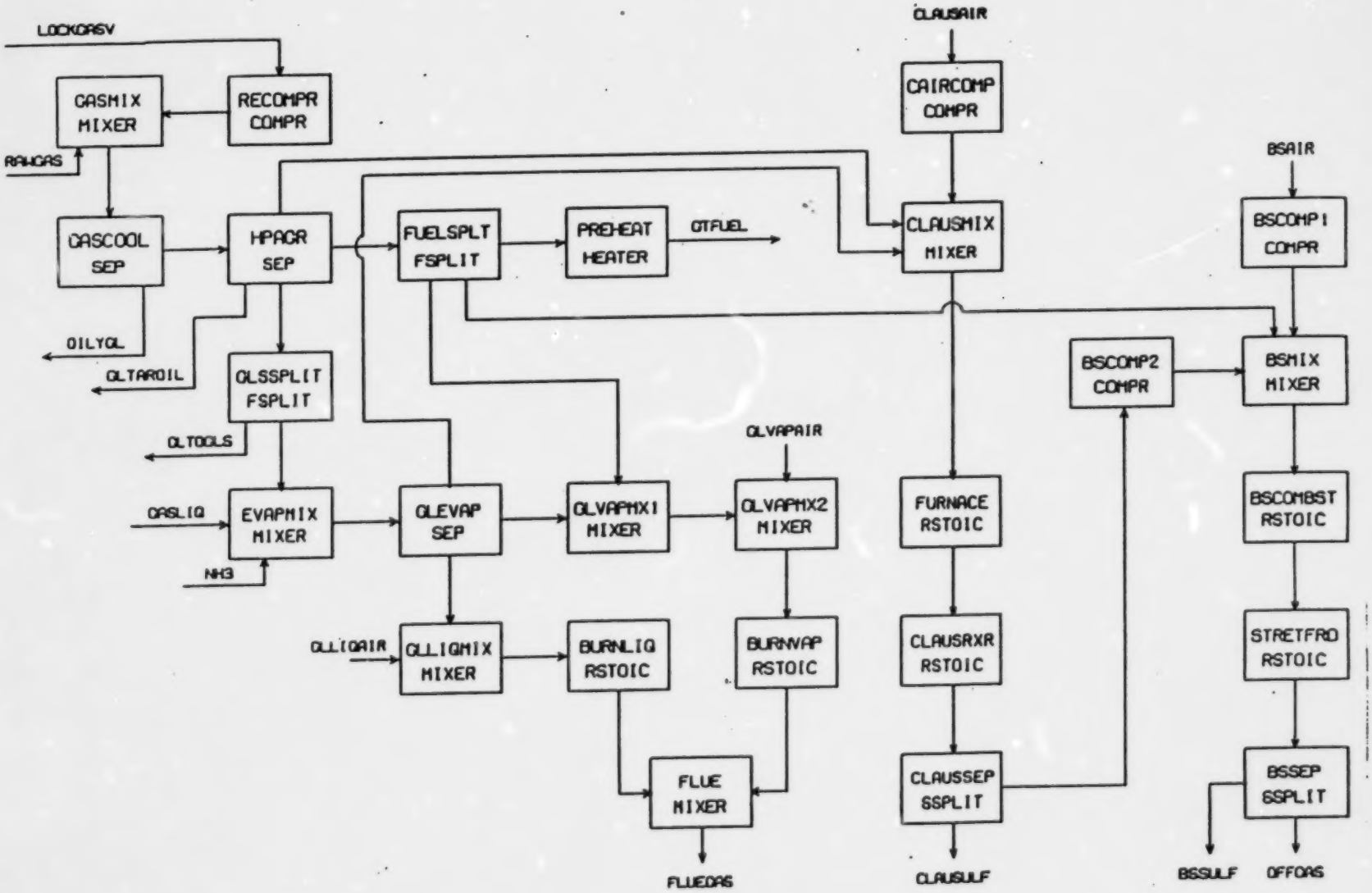


FIGURE 5b. **CLEANUP, LIQWASTE, CLAUS, AND STRETFD FLOWSHEET SECTIONS FROM THE BGC/LURGI SLAGGING FIXED-BED GASIFICATION IGCC PLANT**
(From: Stone (1986))

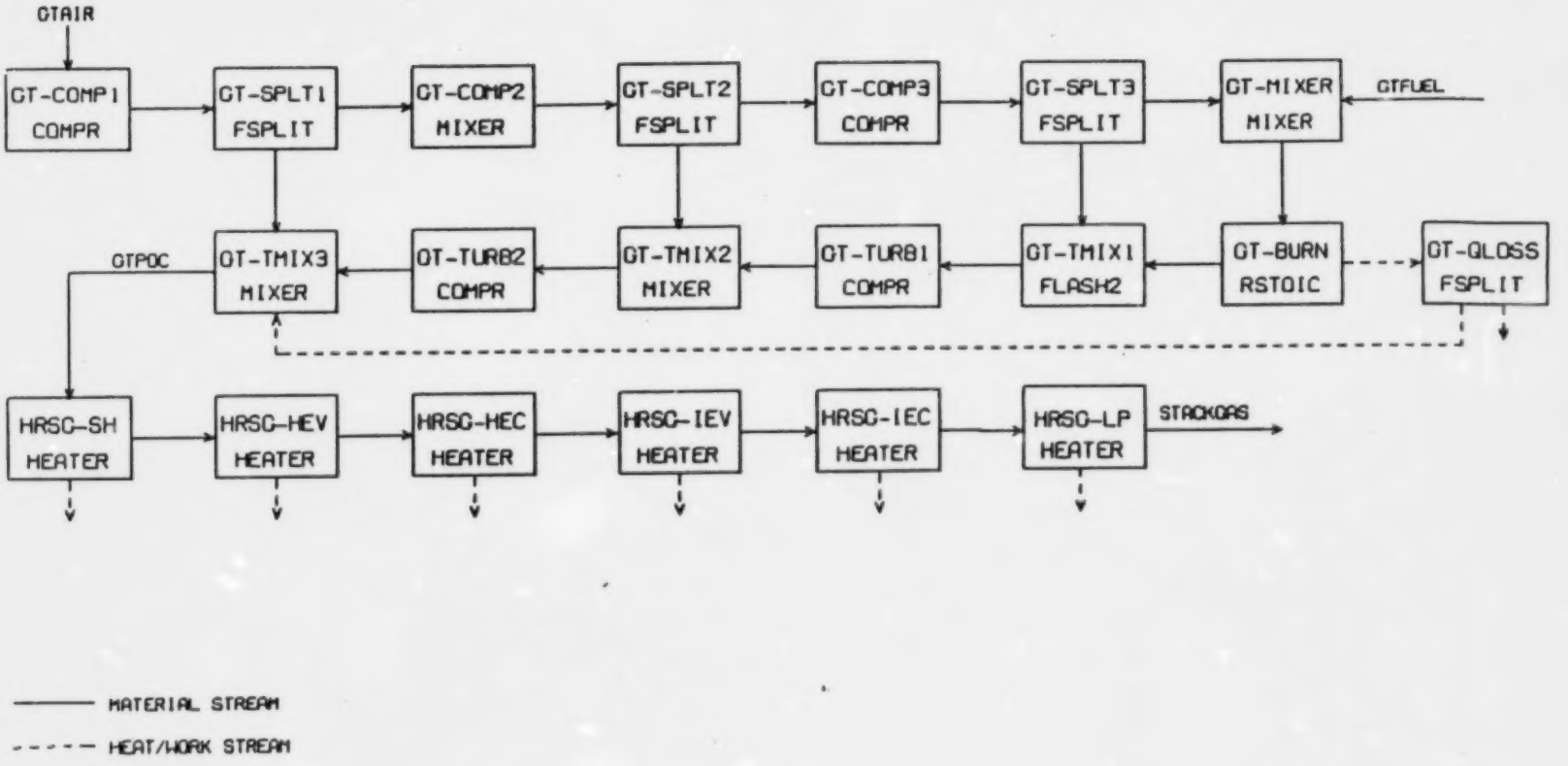


FIGURE 5c. **GASTURB AND HRSG FLOWSHEET SECTIONS FROM THE BGC/LURGI SLAGGING FIXED-BED GASIFICATION IGCC PLANT**
(From: Stone (1986))

COMBUSTION TURBINE GENERATOR SECTION

Assumptions

1. Air compressor, 3 steps: 87% efficient
2. 3% of the flow split for turbine blade cooling at each stage
3. Combustion takes place adiabatically at 2200 °F
4. Gas turbine, 2 stages: 85% efficient

RESULTS

	PFD	SIMULATION
Required air flow (lb/hr)	7,790,335	7,807,700
Power consumed (horsepower)		
Compressor stage 1	---	216,200
Compressor stage 2	---	92,295
Compressor stage 3	---	219,170
Power generated (horsepower)		
Turbine stage 1	---	407,270
Turbine stage 2	---	620,720
Net power (assuming 2% generator loss) (megawatts)	365.35	365.63

TABLE 6a.

TEXACO ENTRAINED FLOW GASIFIER SECTION

Assumptions

1. Decompose coal to elements except 3% of carbon which forms char
2. Burn all O_2 with carbon
3. Burn remaining carbon with CO
4. Burn sulfur with H_2
5. Equilibrium
6. Radiative cooling to 1500 °F
7. Convective cooling to 650 °F
8. Quench ash to 450 °F
9. Generate 1545 psia stream

RESULTS

	PFD	SIMULATION
High Pressure Steam (lb/hr)	1,084,902	1,084,200
Quench water (lb/hr)	-	575,360
Gasifier outlet temp (°F)	-	2101.9
Raw syngas composition (mole 7%)		
H_2	29.72	29.37
CO	38.84	40.02
CO_2	10.60	10.06
CH_4	0.08	0.12
H_2S	0.98	1.03
COS	0.06	.05
N_2	0.71	.73
Ar	0.90	.91
H_2O	18.11	17.67
O_2 --	--	

TABLE 6b.

STEAM TURBINE GENERATOR SECTION

Assumptions

1. High pressure turbine: 85% efficient
2. Intermediate pressure turbine: 90% efficient
3. Medium pressure turbines
   two stages: 90% efficient
4. High pressure BFW pump turbine: 85% efficient

RESULTS

	PFD	SIMULATION
Power generated (horsepower)		
High pressure power turbine	93,000	93,361
Intermediate pressure power turbine	76,770	76,738
Medium pressure power turbine	184,124	170,700
Net power (assuming 2% generator loss) (megawatts)	258.62	248.93
Condenser duty (MM BTU/hr)	1522.65	1557.00
Turbine outlet temperature (°F)		
High pressure power turbine	643.	641.7
Intermediate pressure power turbine	739.	739.6
Medium pressure power turbine	---	141.5

TABLE 6c.

several of the systems. Table 1. shows the model output from ASPEN (USRWEN) by Stephano, (1985) for a KRW-type dry-ash fixed-bed gasifier. Table 2. is output from ASPEN (SLAGGER) modeled by Stone (1986) for a BGC/Lurgi slagging fixed-bed gasification IGCC plant burning Illinous #6 with the following composition:

Proximate	moisture	12.0 %
	Fixed C	54.3 %
	volatiles	35.7 %
	ash	9.12 %
Ultimates	C	69.5 %
	H	5.33 %
	N	1.25 %
	O	10.0 %
	S	3.86 %

This coal is similar in analysis to the Sydney coals (Appendix 3), being high volatile, high sulphur bituminous. This model was for a base-case power plant of 509 MW to produce medium BTU fuel gas for IGCC. The process included Lurgi HP acid gas removal, Claus sulphur recovery, and Beavon-Stretford tailgas treatment. Gas turbines were fired at 2200 ^{o}F. Ambient temperature was assumed as 59^{o}F. The full ASPEN block diagrams are given in Figures 5a - 5c.

A similar data set for Illinois #6 coal is available for the Texaco entrained-bed gasification system. J.S.D. Simulation Service Co. is presently working on an EPRI project to model a nominal 650 MW IGCC power plant using the Texaco system. Fluor Engineers designed a process flow diagram (PFD) based on data from the Cool Water facility. Several sections of the plant (gasifier, combustion turbine, steam turbine generator) have been modeled using ASPEN and results compare very well with Fluor

Tables 6a.-6c). Based on this sucess, EPRI is now modeling the complete IGCC power plant.

BRITISH GAS LURGI: USING SYDNEY COALS

In correspondence with the process developers, it was ascertained that a modified ARACHNE software would be utilized for modeling. Only the gasifier was considered, to arrive at heat and mass balances for the three coals. The gasifier conditions were as follows:

- single 8 foot diameter gasifier
- loading pressure: 450 psig
- coal size range 50mm - 6mm nominal
- hydrocarbons recycled and consumed

A descriptive report for BGC/L is appendixed.

KELLOGG-RUST-WESTINGHOUSE (KRW): USING SYDNEY COALS

As a subcontractor to KRW, the M.W.Kellogg Company conducted the computer simulation using a proprietary KRW gasifier design model. The design was based on the following:

1. KRW air-blown gasification
2. In-bed desulphurization
3. Zinc Ferrite for external desulphurization
4. Hot particulate removal
5. Fuel gas at 1000^{O}F to gas turbine
6. Gas turbine units considered:
   a. Oxygen-rich air with GE MS7001F turbine
   b. air in General Electric 7F turbine

The battery limits of the model begin with the coal feed bin and end at the flange preceeding the gas turbine control valve. The design includes a unit for converting spent in-bed sorbent to a landfillable form meeting U.S. environmental standards.

Three simulations were conducted to predict gasifier performance for the three supplied coal analyses. The plant facilities included are:

1. Coal and limestone preparation
2. Gasification
3. Particulate removal
4. Sulphur removal
5. Heat recovery
6. Ash product oxidation
7. Gas/steam turbines.

The HHV of each coal was calculated by the Dulong Formula. The designs used in the study were not optimized.

TEXACO: USING SYDNEY COALS

Discussions with Mr. Neil Richter (Texaco Montebello Research Laboratory) on June 2, 1989, indicated that Texaco would model only the gasifier. Assumptions for the coal-water slurry feed would be made based on the published information available for Carbogel technology. Results from the Texaco modeling are not expected until August 1989 and will be transmitted when received.

SHELL COAL GASIFICATION PROCESS: USING SYDNEY COALS

Shell researchers claim to use proprietary software developed in-house. Only the gasifier is modeled, not the full IGCC plant, to arrive at mass and heat balances and the suitability of the Shell process for Sydney coals.

Plant sections modeled include:

1. Coal milling and drying

2. Gasification
3. Dry particulate removal
4. Wet particulate removal
5. Sour water treatment

DOW SYNGAS: USING SYDNEY COALS

The Dow researchers used in-house software to model some 15 process streams as identified by the flow diagram (Appendix) to arrive at power balances for a full IGCC system.

DISCUSSION AND CONCLUSIONS

The five systems modeled are substantially different in design which limit comparisons. In summary, the main features are:

BGC/L - fixed-bed, slagging gasifier, lowest temp process oxygen-blown, dry feed

KRW - fluidized-bed, ash agglomerating, med-temp. process air blown, dry feed

DOW - entrained-bed, slagging, high temperature, oxygen-blown, slurry feed

SHELL - entrained-bed, slagging, high temperature, oxygen-blown, slurry feed

TEXACO - entrained-bed, slagging, high temperature, oxygen-blown, slurry feed

SULPHUR REMOVAL

Performance is dependent on the desired output which, in turn, affects the design. For example, only 90-95% sulphur removal is required by Canadian environmental regulations. To meet these specifications, it is not necessary for KRW to achieve 99% sulphur removal as modeled.

Other process developers use a sulphur recovery system such as Claus or Beauron, while KRW utilizes hot gas clean-up. This involves two components:

1. limestone injection into the hot bed like that in a circulating fluid-bed unit; and,
2. zinc-ferrite bed

Limestone injection itself is capable of capturing @ 90% of the sulphur. Therefore, the zinc ferrite bed is not required if only 90% sulphur capture is all that is needed.

This would reduce capital costs and increase overall efficiency of the plant and probably would make KRW look more attractive. The other developers would still have to operate their entire sulphur removal system since this is their single means of removing sulphur rather than having the two-pronged approach of KRW.

LIMESTONE REQUIREMENTS

Table 7 shows the input and out put requirements as pounds per pounds of coal. KRW requires approximately one third as much limestone as coal for Prince or Donkin-Morien, but only half that amount for the Victoria Junction washed coal...likely because of lower sulphur content and the inherent carbonates in the Phalen seam.

The Dow, Shell and Texaco processes do not use limestone for S removal. BGC/L uses some limestone but only as a refluxing agent for caking coals.

OXYGEN, AIR REQUIREMENTS

For the plants designed, the DOW entrained-bed system appears to require more oxygen than the Shell entrained-bed, likely due to the slurry feed as opposed to dry feed for Shell.

KRW is an air-blown system. Therefore, it produces a low-Btu gas (heating value not shown), along with large volumes of

nitrogen.

DOW and Shell are both oxygen-blown gasifiers, they produce a med-Btu gas (heating value not shown for DOW, for Shell = 22,319 GJ/D).

WASTE DISPOSAL

Table 7 shows that KRW generates much more waste than other processes because of the manner of sulphur removal. Thus the disposal of solid wastes will be more costly. Furthermore, the KRW process of limestone injection does not produce the marketable elemental sulphur promised by other desulphurization technology.

FUEL GAS

Fuel gas composition is not available for all the processes (Table 8). Untreated DOW and Shell fuel gas will be "sour" with an estimated 1 to 2% H_2S and dominantly carbon monoxide (60-70%). The Shell process results in a high hydrogen gas, while DOW predicts high carbon dioxide.

Treated fuel gas (i.e. desulphurization) will result in a 98.5 to 99% reduction in H_2S. The treated KRW gas is dominantly nitrogen and carbon dioxide since it is derived from an air-blown system. The Texaco process yields a gas of much higher moisture content because of the slurry feed. Otherwise, the composition is high in hydrogen, carbon monoxide and carbon dioxide.

PLANT POWER, EFFICIENCY

Table 9 summarizes the computer simulation for plant power, heat rate, and efficiency. The total plant efficiencies estimated by DOW and KRW range from 40.8 to 43.6%. Such percentages are very high in view of other reports (e.g. Stone 1986 at 39%, Table 2) and puts the performance off the scale of Figure 4b. We consider these results to be overly optimistic by the process developers.

Plant efficiencies appear to decrease in the sequence of coals: V>D>P and Prince coal requires the greatest coal feed for net electric output.

TABLE 7. COAL GASIFIER: SUMMARY FROM COMPUTER MODELS

PROCESS Parameter (lb/lb coal)	COAL	BGC/L	KRW	DOW	SHELL	TEXACO
Oxygen to gasifier	P	0.640	0.000	0.904	0.851	p
	D	0.645	0.000	0.818	na	p
	V	0.672	0.000	0.824	na	p
Air to Coal prep & gasifier	P	0.000	0.323	0.000	0.185	p
	D	0.000	0.291	0.000	na	p
	V	0.000	0.288	0.000	na	p
Limestone	P	0.000	0.368	0.000	0.000	p
	D	0.000	0.353	0.000	0.000	p
	V	0.000	0.172	0.000	0.000	p
Solid Waste	P	0.168	0.449	0.160	0.140	p
	D	0.094	0.383	0.104	na	p
	V	0.065	0.204	0.067	na	p
Elemental Sulphur	P	0.050	0.000	0.039	0.039	p
	D	0.041	0.000	0.035	0.035	p
	V	0.022	0.000	0.019	0.019	p

NOTE: Data based on total coal input as fed.
P= Prince; D= Donkin, V= VJ mix, p= pending

TABLE 8. FUEL GAS COMPOSITION: SUMMARY FROM COMPUTER SIMULATION.

A. UNTREATED FUEL GAS

PROCESS:	B.G.C./Lurgi			KRW			DOW			SHELL			TEXACO		
GAS: COAL:	P	D	V	P	D	V	P	D	V	P	D	V	P	D	V
H2	28.66	27.02	27.87	na	na	na	3.51	3.71	3.89	30.60	na	na	p	p	p
CO	57.00	58.99	59.54	na	na	na	61.86	69.47	71.45	61.60	na	na	p	p	p
CO2	1.99	1.80	1.37	na	na	na	28.08	20.61	19.19	1.50	na	na	p	p	p
H2S	1.46	1.23	0.64	na	na	na	2.12	1.82	0.96	1.30	na	na	p	p	p
COS	0.08	0.07	0.04	na	na	na	0.17	0.14	0.07	0.20	na	na	p	p	p
CH4 **	7.97	7.87	7.55	na	na	na	0.27	0.54	0.65	0.01	na	na	p	p	p
N2 *	2.83	2.82	2.82	na	na	na	3.71	3.43	3.48	3.40	na	na	p	p	p
AR	na	na	na	na	na	na	na	na	na	1.10	na	na	p	p	p
H2O	na	na	na	na	na	na	0.27	0.30	0.31	0.30	na	na	p	p	p
NH3	na	na	na	na	na	na	0.00	0.00	0.00	0.00	na	na	p	p	p
HCN	na	na	na	na	na	na	0.01	0.00	0.01	0.00	na	na	p	p	p
TOTAL	99.99	99.80	99.83	na	na	na	100.0	100.0	100.0	100.0	na	na	p	p	p

B. TREATED FUEL GAS

PROCESS:	B.G.C./Lurgi			KRW			DOW			SHELL			TEXACO		
GAS: COAL:	P	D	V	P	D	V	P	D	V	P	D	V	P	D	V
H2	na	na	na	1.84	1.70	1.88	na	na	na	31.30	na	na	p	p	p
CO	na	na	na	10.97	10.55	11.25	na	na	na	62.80	na	na	p	p	p
CO2	na	na	na	26.29	26.38	25.79	na	na	na	1.00	na	na	p	p	p
H2S	na	na	na	0.00	0.00	0.00	na	na	na	0.02	na	na	p	p	p
COS	na	na	na	0.00	0.00	0.00	na	na	na	0.01	na	na	p	p	p
CH4	na	na	na	0.91	0.86	0.83	na	na	na	0.01	na	na	p	p	p
N2 *	na	na	na	43.35	44.51	44.00	na	na	na	3.40	na	na	p	p	p
AR	na	na	na	0.00	0.00	0.00	na	na	na	1.10	na	na	p	p	p
H2O	na	na	na	16.62	15.98	16.23	na	na	na	0.40	na	na	p	p	p
NH3	na	na	na	0.01	0.01	0.01	na	na	na	0.00	na	na	p	p	p
HCN	na	na	na	0.00	0.00	0.00	na	na	na	0.00	na	na	p	p	p
TOTAL	na	na	na	99.99	99.99	99.99	na	na	na	100.0	na	na	p	p	p

NOTES: * Volume for N2 includes argon for Dow. na= not analysed. p= data pending.
** Includes C2H6, C2H4

TABLE 9. COMPUTER SIMULATION: POWER, HEAT RATE, EFFICIENCY

PROCESS: COAL		BGC/L	DOW	KRW	SHELL	TEXACO	
PARAMETER:							
COAL FEED	P	643	1797.6	2005.4	1000.0		p
(as received	D	611	1582.8	1988.4	na		p
tons/day)	V	605	1476.0	1792.7	na		p
GROSS ELECTRIC	P	na	235.4	2654.0	na		p
(output in MW)	D	na	229.1	268.9	na		p
	V	na	227.8	261.0	na		p
NET ELECTRIC	P	na	211.9	252.8	na		p
(output in MW)	D	na	209.8	255.8	na		p
	V	na	209.8	249.7	na		p
PLANT HEAT RATE	P	na	8278.0	8477.0	na		p
(BTU/kWh	D	na	7987.0	8526.0	na		p
for HHV)	V	na	7836.0	8352.0	na		p
PLANT EFFICIENCY	P	na	41.2	40.3	na		p
(%)	D	na	42.7	40.8	na		p
	V	na	43.6	41.2	na		p

LEGEND: P= Prince Mine, Hub seam; D= Donkin-Morien, Harbour seam; V= VJ Wash Plant, thermal, Phalen & Harbour.
p= results pending. na= not analysed.

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APPENDIX 1.

MATHEMATICAL MODEL FOR IGCC POWER PLANT

SIMPLIFIED ANALYTICAL IGCC SYSTEM MODEL

The following is a simplified energy flow model used to develop a system efficiency relationship for a coal-gasifier combined cycle. (After: Stochl and Nainiger, 1983):

SYMBOLS

a	fraction of raw-gas chemical energy that enters gas turbine
b	fraction of raw-gas chemical energy lost in cleanup systems
f	fraction of Q_{exh} transferred to steam cycle
K^i	influence coefficient for p^i
p^i	parameter i
P_{aux}	auxiliary power requirements
P_{gt}	gas turbine power
P_{net}	net system power
P_{st}	steam turbine power
Q_c	coal energy input based on higher heating valve
Q_{ch}	chemical energy in raw gas
Q_{exh}	energy content of gas-turbine exhaust
Q_g	energy input to steam cycle from gasifier
Q_{gt}	fuel energy to tas turbine, $aQ_{ch} + Q_{th}$
Q_m	energy input to steam cycle from the cooling of compressor discharge air used for turbine cooling
Q_{rg}	energy content of raw gas = $Q_{ch} + Q_{th}$
Q_{st}	steam-turbine energy, = $xQ_{th} + fQ_{exh} + Q_g + Q_m$
Q_{th}	thermal energy in raw gas
x	fraction of raw-gas thermal energy directly transferred to steam cycle
y	fraction of raw-gas thermal energy directly transferred to gas turbine
z	fraction of raw-gas thermal energy that does not directly enter either gas turbine or steam turbine
n_{cc}	net IGCC system efficiency
n_{gt}	net gas-turbine efficiency
n_{st}	net steam-turbine efficiency

We may define net IGCC system efficiency as

$$\eta_{cc} = \frac{P_{net}}{Q_c} = \frac{P_{gt}}{Q_c} + \frac{P_{st}}{Q_c} - \frac{P_{aux}}{Q_c}$$

$$= \left(\frac{P_{gt}}{Q_{gt}}\frac{Q_{gt}}{Q_c}\right) + \frac{P_{st}}{Q_{st}}\frac{Q_{st}}{Q_c} - \frac{P_{aux}}{Q_c}$$

$$= \eta_{gt}\frac{Q_{gt}}{Q_c} + \eta_{st}\frac{Q_{st}}{Q_c} - \frac{P_{aux}}{Q_c}$$

where

$$Q_{gt} = aQ_{ch} + yQ_{th}$$

$$Q_{st} = xQ_{th} + fQ_{exh} + (Q_m + Q_{gt})$$

Substituting yields

$$\eta_{cc} = \eta_{gt}\left(a\frac{Q_{ch}}{Q_c} + y\frac{Q_{th}}{Q_c}\right) + \eta_{st}\left(x\frac{Q_{th}}{Q_c} + f\frac{Q_{exh}}{Q_c} + \frac{(Q_m + Q_g)}{Q_c}\right) - \frac{P_{aux}}{Q_c}$$

But

$$\frac{Q_{exh}}{Q_c} = \frac{Q_{exh}}{Q_{gt}}\frac{Q_{gt}}{Q_c} = \frac{Q_{exh}}{Q_{gt}}\left(a\frac{Q_{ch}}{Q_c} + y\frac{Q_{th}}{Q_c}\right)$$

so

$$\eta_{cc} = \left(\eta_{gt} + \eta_{st}\frac{Q_{exh}}{Q_{gt}}\right)a\frac{Q_{ch}}{Q_c}$$

$$+ \left[x\eta_{st} + y\left(\eta_{gt} + f\eta_{st}\frac{Q_{exh}}{Q_{gt}}\right)\right]\frac{Q_{th}}{Q_c} + \left(\frac{Q_m + Q_g}{Q_c}\right)\eta_{st} - \frac{P_{aux}}{Q_c} \qquad \text{(C1)}$$

General Influence Coefficients

Equation (C1) indicates that the net IGCC system efficiency is a function of 11 parameters:

$$\eta_{cc} = \eta_{cc}\left(a, x, y, z, f, \frac{Q_{ch}}{Q_c}, \frac{Q_{th}}{Q_c}, \eta_{gt}, \eta_{st}, \frac{Q_m + Q_g}{Q_c}, \frac{P_{aux}}{Q_c}\right)$$

A general influence coefficient can be defined as being the relative change in net system efficiency caused by a relative change in one of the 11 parameters; or

$$\frac{\partial \eta_{cc}}{\eta_{cc}} = K^i \frac{\partial p^i}{p^i}$$

where K^i, the influence coefficient for p^i, is defined as

$$K^i = \frac{\partial \eta_{cc}}{\partial p^i} \frac{p^i}{\eta_{cc}}$$

Coefficient K^i can be evaluated from equation (C1) by differentiating with respect to one of the 11 parameters and holding the remaining 10 parameters constant. In order to obtain the influence coefficient for η_{gt} and x, equation (C1) will be modified by assuming that $Q_{exh}/Q_{gt} = 1 - (\eta_{gt} + C_1)$, where C_1 is a constant that includes gas-turbine combustor, generator, and mechanical losses, and by making use of the relation $x + y + z = 1$ and assuming that z = constant.

In an IGCC system using a particular gasifier a, Q_{ch}/Q_c and Q_{th}/Q_c would be constant parameters. The more significant influence coefficients for the IGCC systems considered in this analysis are

$$K_{gt} = \frac{\partial \eta_{cc}/\eta_{cc}}{\partial \eta_{gt}/\eta_{gt}} = \left(1 - f\eta_{st}\right)\left(a\frac{Q_{ch}}{Q_c} + y\frac{Q_{th}}{Q_c}\right)\frac{\eta_{gt}}{\eta_{cc}} \qquad (C2)$$

$$K_{st} = \frac{\partial \eta_{cc}/\eta_{cc}}{\partial \eta_{st}/\eta_{st}} = \left[\left(a\frac{Q_{ch}}{Q_c} + y\frac{Q_{th}}{Q_c}\right) f\frac{Q_{exh}}{Q_{gt}} + x\frac{Q_{th}}{Q_c} + \frac{Q_m + Q_g}{Q_c}\right]\frac{\eta_{st}}{\eta_{cc}} \qquad (C3)$$

$$K_y = \frac{\partial \eta_{cc}/\eta_{cc}}{\partial y/y} = \left[\eta_{gt} - \eta_{st}\left(1 - f\frac{Q_{exh}}{Q_{gt}}\right)\right]\frac{Q_{th}}{Q_c}\frac{y}{\eta_{cc}} \qquad (C4)$$

$$K_f = \frac{\partial \eta_{cc}/\eta_{cc}}{\partial f/f} = \eta_{st}\frac{Q_{exh}}{Q_{gt}}\left(a\frac{Q_{ch}}{Q_c} + y\frac{Q_{th}}{Q_c}\right)\frac{f}{\eta_{cc}} \qquad (C5)$$

$$K_{(Q_m+Q_g)/Q_c} = \frac{\partial \eta_{cc}/\eta_{cc}}{\dfrac{\partial\left[(Q_m + Q_g)/Q_c\right]}{(Q_m + Q_g)/Q_c}} = \frac{\eta_{st}\left[(Q_{ch} + Q_g)/Q_c\right]}{\eta_{cc}} \qquad (C6)$$

$$K_{P_{aux}/Q_c} = \frac{\partial \eta_{cc}/\eta_{cc}}{\dfrac{\partial(P_{aux}/Q_c)}{P_{aux}/Q_c}} = -\frac{P_{aux}/Q_c}{\eta_{cc}} \qquad (C7)$$

APPENDIX 2.

COAL DATA REQUIREMENTS FOR MODELING

BRITISH GAS CORPORATION

Request for Slagging Gasifier Performance Prediction

Coal Characterisation Data

This document provides a format for the presentation of basic coal characterisation data in the form of physical and chemical properties of the coal. The tests used to obtain this information should be taken from British Standards Methods for sampling and analysis of coal, or be methods which have been shown to be equivalent.

The results are to be reported either as analysed (ad) or on the dry basis (db), methods for converting to bases other than these can be found in BS1016 : Part 16, 1971.

NOTE: The essential items of information which must be provided are: Proximate and Ultimate Analysis, Ash Analysis, Crucible Swelling No. or Gray-king Coke Type, Calorific Value, weight percent of coal smaller than ¼ in. (6mm) as delivered to site.

A. CHEMICAL DATA

1.

Proximate Analysis (% w/w)		
1. Moisture	(a.d.)	
2. Ash	(d.b.)	
3. Volatile matter	(d.b.)	
4. Fixed carbon	(d.b.)	

2.	Ultimate Analysis (% d.b. w/w)	
	1. Carbon	
	2. Hydrogen	
	3. Nitrogen	
	4. Oxygen (a) determined	
	(b) calculated (by difference)	
	5. Sulphur (total)	
	6. Sulphur (organic)	
	7. Sulphur (pyritic)	
	8. Sulphur (sulphate)	
	9. Chlorine	
	10. Carbon dioxide	
	11. Phosphorus	

3.	Gross Calorific Value (db) (Btu/lb)	

4.

Ash Analysis (% w/w)	
as oxide	as element
1. SiO_2	Si
2. Al_2O_3	Al
3. Fe_2O_3	Fe
4. CaO	Ca
5. MgO	Mg
6. Na_2O	Na
7. K_2O	K
8. TiO_2	Ti
9. Mn_3O_4	Mn
10. SO_3	S
11. P_2O_5	P
Silica Number	

C. PHYSICAL DATA

1. Dilatometry

1. Softening temperature (°C), T1	
2. Max. contraction temp (°C), T2	
3. Max dilatation temp. (°C), T3	
4. Contraction, C, (%)	
5. Dilatation, D, (%)	
6. (C+D) (%)	

2.

Crucible Swelling Number	

3. Ash Fusion Characteristics (°C)

Conditions :	Oxidising	Reducing
1. Deformation temp.		
2. Hemispherical temp.		
3. Flow temp.		

D. ADDITIONAL INFORMATION

1. Gray-King Coke type

2. Hardgrove Index

3. Coal size Distribution (wt%)

Above 2"
2" - 1¼"
1" - 1¼"
¾" - 1"
½" - ¾"
3/8"- ½"
¼" - 3/8"
1/8"- ¼"
below 1/8"

(or equivalent metric dimensions: 50, 35, 25, 18, 12, 10, 6, 3mm)

BRITISH GAS : WESTFIELD DEVELOPMENT CENTRE

TYPICAL COAL ANALYSES

	PHASE II			PHASE II	PHASE III	PHASE II	PHASE III
	Seafield	Manton	Ohio 9	Pittsburgh	Pittsburgh	Markham	Markham
Proximate (wt %)	As Received	As Received	As Received	As Received	As Received	As Received	As Received
Moisture	12.01	4.97	2.75	4.21	6.81	8.38	12.29
Ash	19.67	7.14	10.06	10.68	14.43	4.75	5.67
Fixed Carbon	41.81	55.17	47.87	50.90	45.28	56.18	50.85
Volatile Matter	26.51	32.72	39.32	34.21	33.48	30.69	31.19
Ultimate (wt %)	Air Dried	Air Dried	Air Dried	Air Dried	Air Dried	Air Dried	Air Dried
Carbon	58.01	76.29	68.30	72.50	68.60	72.32	73.32
Hydrogen	3.97	4.58	5.10	4.93	4.56	4.27	4.63
Oxygen	8.55	4.74	8.13	5.89	5.20	7.13	5.30
Nitrogen	0.98	1.45	0.80	1.43	1.46	1.37	1.57
Sulphur	0.48	2.03	3.59	1.85	2.92	1.42	1.64
Ash Analysis (wt %)							
SiO_2	51.98	41.72	46.42	51.79	46.0	34.93	34.80
Al_2O_3	29.10	26.10	22.09	25.22	20.8	24.83	27.60
Fe_2O_3	4.69	20.41	21.65	15.86	15.4	22.59	18.70
TiO_2	N/D	N/D	N/D	N/D	1.1	N/D	1.20
CaO	4.22	2.16	1.53	1.56	5.2	4.59	3.30
MgO	2.59	0.92	1.07	0.76	0.83	1.43	0.98
SO_3	N/D	N/D	N/D	N/D	N/D	N/D	N/D
K_2O	1.00	2.16	2.53	2.05	1.82	1.72	1.87
Na_2O	0.39	1.83	0.48	0.38	0.46	3.36	2.47
P_2O_5	N/D	N/D	N/D	N/D	N/D	N/D	N/D
Unknown	N/D	N/D	N/D	N/D	N/D	N/D	N/D
Heating Value							
HHV (Btu/lb) Air Dried	10,197	13,469	12,432	12,939	12,446	12,466	13,235
HHV (Btu/lb) Dry	10,879	13,888	12,619	13,213	12,710	13,334	13,853
Swelling No.	1	-	3.5 - 6	7	6	½ - 1	½ - 1
Gray King	A	-	G2 - G4	G6	G7	C - D	C - D
% Gasification Efficiency (exc. steam & oxygen)	83.94	89.51	83.32	89.18	87.44	88.59	92.34
% Efficiency (incl. steam & oxygen estimated requirements)	72.56	77.70	72.51	76.79	77.25	78.22	81.25

SHELL COAL GASIFICATION PROCESS

The following analytical date was requested in order to conduct a simulation of the gasifier:

Proximate analysis (wt % as received):

Moisture	-
Ash	-
Volatile matter	-
Fixed carbon	-

Elemental analysis (wt % as received):

Carbon	-
Hydrogen	-
Nitrogen	-
Sulphur	-
Ash	-

Lower Heating Value (As received):

Hardgrove index

Ash fusibility data (Reducing conditions):

- Initial deformation temperature
- Softening temperature
- Hemispherical temperature
- Fluid temperature

Ash composition:

DOW CHEMICAL

Coal analytical information requested for computer simulation:

Coal Data

- Ultimate Analysis
- Proximate Analysis
- Ash Analysis
- Heating Value
- Equilibrium Moisture Content

Environmental Standards

- Sulfur Removal
- NOx Emissions

Combustion Turbine

- Options are:
  - Westinghouse W501D5
  - General Electric 7E, 7F
  - Others

Misc. Bases:

- Ambient Air Temp
- Atmospheric Pressure
- Cooling System

APPENDIX 3.

INFORMATION SUPPLIED TO PROCESS DEVELOPERS FOR MODELING OF SYDNEY COALS

COAL DATA AND IGCC POWER PLANT DESIGN CRITERIA

Coal Analysis for:

1. Prince Colliery Hub seam
2. Donkin-Morien Harbour seam
3. Victoria Junction (VJ) thermal coal

Data to be supplied:

Ultimate Analysis
Proximate Analysis
Ash Analysis
Heating Value
Moisture content
Ash Fusion Temperatures
Hardgrove Grindability Index
Free Swelling Index
(see attached pages)

Environmental Standards:

1. Sulphur removal - 98%
2. NOx emmisions - Maximum fuel gas moisturization to minimize NOx formation

Combustion Turbine:

- Advance GE MS7001F

IGCC Power Plant Specifications:

- Nominal plant capacity:200MW
- Cycling capability down to 50%
- Cooling System: once through, seawater cooling
- Inlet cooling water temperature rante: -1°C to +15°C
- Allowed cooling water temperature rise: 10°C
- IGCC power plant is to be designed for an ambient air temperature of 0°C (32°F) and an atmospheric pressure of 100.54 kPa.

Cooling Water Temperature:

- Inlet temperature range -1°C to $^{+}$5°C
- Allowed temperature rise: 10°C

COAL ANALYSIS

Mine	Prince Hub		Donkin (Tunnel) Harbour		Phalen/Lingan Phalen/Harbour	
Seam						
	(As Received)	(Dry)	(As Received)	(Dry)	(As Received)	(Dry)
Proximate Analysis						
Moisture	6.71%	0.00%	6.50%	0.0 %	8.50%	0.0 %
Ash	11.61%	12.45%	7.48%	8.0 %	4.67%	5.1 %
Volatile Matter	33.02%	35.40%	33.66%	36.0 %	32.67%	35.7 %
Fixed Carbon	48.66%	52.15%	52.36%	56.0 %	54.16%	59.2 %
Total	100.00%	100.00%	100.00%	100.0 %	100.00%	100.0 %
Gross Heating Value						
KJ/Kg	27,203	29,159	29,578	31,634	31,072	33,959
(Btu/lb)	(11,695)	(12,536)	(12,716)	(13,600)	(13,358)	(14,600)
Ultimate Analysis						
Carbon	66.78%	71.58%	71.06%	76.0 %	74.11%	81.0 %
Hydrogen[1]	4.94%	5.29%	4.68%	5.0 %	4.76%	5.2 %
Nitrogen	1.36%	1.46%	1.21%	1.3 %	1.37%	1.5 %
Ash	11.61%	12.45%	7.48%	8.0 %	4.67%	5.1 %
Sulphur	4.07%	4.36%	3.55%	3.8 %	1.92%	2.1 %
Oxygen[1]	4.53%	4.86%	5.52%	5.9 %	4.67%	5.1 %
Total	93.29%	100.00%	93.50%	100.0 %	91.50%	100.0 %
Chlorine (dry basis)		0.30%		0.15%		0.23%

Note: 1. Hydrogen and Oxygen in sample moisture are not included

	Prince Hub	Donkin-Morien Harbour	VJ Thermal
Ash Analysis			
SiO_2	41.16%	33.09%	35.84%
Al_2O_3	16.94%	19.23%	18.49%
TiO_2	0.60%	0.70%	0.87%
Fe_2O_3	34.10%	35.30%	34.23%
CaO	1.77%	5.15%	3.55%
MgO	0.86%	0.54%	0.82%
Na_2O	0.70%	0.75%	0.72%
K_2O	1.60%	0.98%	1.56%
P_2O_5	0.15%	0.35%	0.55%
SO_3	2.02%	3.81%	3.27%
MnO_2	0.10%	0.10%	0.10%
Total	100.00%	100.00%	100.00%

Ash Fusion Temp, °C	Red	(Oxid)	Red	(Oxid)	Red	(Oxid)
IT	1055	(1345)	1070	1350	1082	1330
ST	1095	(1390)	1125	1400	1120	1400
HT	1180	(1404)	1210	1410	1165	1410
FT	1305	(1410)	1270	1460	1305	1420
Hardgrove Grindability Index	56		60		57	
Free Swelling Index	5.5(AVG)-6.5(MAX)		7.5		7.5	

APPENDIX 4.

PROCESS DEVELOPER REPORTS

International Consultancy Service
a part of British Gas plc

EVALUATION OF
CANADIAN COALS AS FEEDSTOCKS FOR
THE BGL GASIFIER

For Atlantic Coal Institute

July 1989

CONFIDENTIAL

British Gas

BRITISH GAS plc

Performance Predictions of the BGL gasifier for Atlantic Coal Institute

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2. Features of the BGL Gasification Process

3. Description of a Typical BGL - GCC Plant

4. Conclusions

Appendix 1	Performance prediction of the BGL Gasifier on Prince Colliery Hub Coal
Appendix 2	Performance prediction of the BGL Gasifier on Donkin-Morien Harbour Coal
Appendix 3	Performance prediction of the BGL Gasifier on Phalen Harbour coal.
Figure 1	The British Gas Lurgi (BGL) Slagging Gasifier
Figure 2	BGL - Gasification Combined Cycle Base Scheme

BRITISH GAS plc

Performance Predictions of the BGL gasifier for Atlantic Coal Institute

1.0 INTRODUCTION

This report presents the results of computer simulations of the performance of the British Gas / Lurgi (BGL) slagging gasifier on three coals : Prince Colliery Hub seam, Donkin-Morien Harbour seam and Phalen Harbour coal. The simulations were undertaken in response to a request from Atlantic Coal Institute for such data. It is intention of ACI to use the data provided by British Gas in this report to model IGCC plants based on the BGL gasifier.

The results of the computer simulations for each of the coals mentioned above are given in Appendices 1 to 3. These comprise an overall energy and material balance over a single 8 foot nominal diameter gasifier operating at its standard loading and at a pressure of 450 psig. The simulations are based on the assumption that the feed coal is made available in a specified size range (50mm - 6mm nominal) and that all the condensible hydrocarbons produced during gasification of the coals are recycled and consumed within the gasifier.

2.0 FEATURES OF THE BGL GASIFICATION PROCESS

The BGL gasifier (Figure 1) is a state of the art, second generation, fixed bed gasifier in which coal is gasified under pressure using steam and oxygen. It utilises a simple water-cooled double-wall reactor based on the design of the Lurgi gasifier which has seen widespread commercial operation over several decades. Coal is fed into the top of the gasifier via a simple, well-proven loading system and descends countercurrent to the products of gasification. Steam and oxygen are injected into the hearth of the gasifier via tuyeres which are comparable with those used in blast furnace practice.

The countercurrent nature of the process ensures that this type of gasifier displays a very high thermal efficiency, exceeding that of entrained or fluidised bed systems. The total ash content of the coal is discharged from the gasifier as an odourless, non-leachable, carbon-free, inert, granular, glassy frit and is suitable for landfill. The slag tapping system is fully automatic and readily responds to variations in slag quantities, composition and viscosity. Liquid hydrocarbon byproducts produced are readily separated from the product gas and recycled via a closed loop circuit to the gasifier for gasification to extinction.

BGL gasification has a net overall cold gas efficiency for the conversion of coal to clean fuel gas in excess of 90% when gasifying a typical bituminous coal such as those used for the bulk of power generation from coal in the UK. This is much higher than for competing gasification technologies. Carbon conversion to product gas is well in excess of 99%.

Among the attractive features of the BGL gasification system are the following:

1. The high thermal efficiency of the fixed bed gasification system leads to a low oxygen requirement.

2. The presence of a substantial amount of coal in the gasifier bed allows continued gasification, even if the coal supply is temporarily halted and consequently enhanced safety during operation.

3. The throughput of the gasifier can be varied rapidly with no control problems. The gasifier is readily put on standby and can be rapidly returned to gasification allowing for intermittent operation of the gasifier when required by product gas needs.

4. Substantial and unpredictable variation in coal composition ash content and ash analysis can be tolerated.

5. Due to the gasifier's countercurrent operation the high temperature zone is restricted to the area in front of the tuyeres where the oxygen input is consumed and hence refractory life is very extensive. Endothermic steam reactions above this combustion zone lead to a rapid decrease in bed temperatures as the gasification products ascend the gasifier.

6. The outlet temperature of the product gases is low with a relatively low sensible heat content. Typically this temperature is 400-500°C, leaving only a modest duty for downstream heat recovery equipment.

7. The tuyere system allows substantial quantities of coal fines and all the tar and oil byproducts to be injected into the gasifier for gasification or incineration. The coal fines can be injected either pneumatically using a carrier gas or as a coarse coal-water slurry.

8. The materials of construction requirements for the gasifier and the associated downstream equipment can be met by simple materials, mainly carbon steel. Equipment can be readily shop fabricated and moved to site.

9. A wide variation in coal types can be gasified regardless of swelling and caking properties.

10. The calorific value of the product gas is comparatively high due to the significant production of methane during gasification particularly from the pyrolysis zone. This makes the BGL product gas very suitable for the production of SNG by methanation at high efficiency. Alternatively, the methane in the raw gas can be separated out for use as SNG while the remaining gas consisting mainly of hydrogen and carbon monoxide can be used to fuel power generation plant.

11. The use of the BGL gasifier enables coal to be utilised with minimal environmental impact.

3.0 DESCRIPTION OF A TYPICAL BGL - GCC PLANT

This type of power plant comprises the following units (Figure 2).

a. COAL GASIFICATION plant for the production of clean fuel gas and for the treatment of process water and off gases. The latter includes both sulphur removal and recovery of saleable sulphur.

b. COMBINED CYCLE plant for electric power generation.

c. OFFSITES PLANT include air separation plant for the production of oxygen, boiler feed water treatment unit, coal/slag storage and handling facilities.

3.1 Fuel Gas Production and Treatment

The coal is transported from the coal stockpile to the BGL gasifiers where it is converted into a medium-Btu gas using steam and oxygen as gasification agents. The product gas is quenched and saturated with water and fed at about 160°C to the low temperature gas cooling where BFW is preheated. The cooled raw gas is routed to the gas purification system for sulphur removal.

The sour gases are sent to an oxygen Claus plant which recovers saleable sulphur and produces a tail gas containing CO and H_2 and residual sulphur compounds (H_2S) which is recycled to the crude fuel gas. With this concept sulphur recovery efficiencies greater than 99% can be achieved. Another merit of the concept is that no off gas is released to the atmosphere.

The purified fuel gas is routed to the gas turbines without any further treatment.

The process waste water streams are collected and fed to a thermal oxidation stage. Thus, the gasification plant does not produce any effluents. The only waste product to be disposed of is the chloride salt from the coal, which is discharged as a solid residue. Alternatively, a conventional, biological treatment process can be used which produces a water stream suitable for process re-use.

3.2 Combined Cycle Plant

The gas turbine makes optimum use of fuel gas, ie no heat exchange is required between the combustion process and the turbine. On the other hand, waste heat from the steam turbine condenser can be discharged virtually at ambient temperature. This leads to very high thermal efficiencies.

4.0 CONCLUSIONS

The three coals evaluated (Prince Colliery Hub seam, Donkin-Morien Harbour seam and Phalen Harbour coal) are all well-suited to BGL gasification and yield cold gas efficiencies of better than 90% on simulation of the gasification process. However, the overall efficiency of an IGCC plant based on the BGL gasifier, is critically dependent upon the integration of the various processing stages in the plant. BGL have considerable expertise in undertaking such studies which can be offered under normal commercial terms.

BRITISH GAS plc

Performance Predictions of the BGL gasifier
for Atlantic Coal Institute

Appendix 1

Performance prediction of the BGL Gasifier on Prince Colliery Hub Coal

Notes

1. Gasifier Loading = 960 lb (DAF)/sq ft h

2. Cold Clean Gas Efficiency = 91.85% The cold clean gas efficiency is defined as the percentage of the potential heat of the feed coal in the cold clean (excluding ammonia, hydrogen chloride, hydrogen cyanide and sulphur compounds) gas product from the gasifier.

3. The Institute of Gas Technology (IGT), Chicago, correlation was used to calculate the coal HHV since the quoted CV gave an unrealistically low offtake temperature for the gasifier.

4. Desulphurised product gas was assumed for the burner gas with air as the oxidant. The burner, which is part of the slag tapping system, is located below the slag tap and ensures the maintenance of correct slag tapping temperatures. It is also used for start-up.

DATA SUMMARY SHEET
ATLANTIC COAL INSTITUTE ENQUIRY.

British Gas plc
Midlands Research Station

Coal Type : PRINCE COAL.

Coal Analysis

	Ultimate dry ash free wt %		Proximate as received wt %
C	81.48	Fixed Carbon	48.66
H	6.02	Volatile Matter	33.02
O	5.53	Moisture	6.71
N	1.66	Ash	11.61
S	4.97		
Cl	0.34		
	100.00		100.00

Coal feedrate	(as received)	643 tonne/d
Coal feedrate	(dry ash free)	21888 kg/h
Gasifier loading	(dry ash free)	4687 kg/h m²
No of gasifiers	(excludes spares)	1 x 2.44 m diameter
Gasification pressure		32.0 bar
Steam feed pressure		38.9 bar
Oxygen feed pressure		37.6 bar

Gasifier Performance

	dry gas ex gasifier
Gas output, million MJ/d #	16.890
Gas output, million Nm³/d #	1.162
Gas HHV, MJ/Nm³ #	14.54
Outlet temperature, °C	465

	kg/kg coal (dry ash free)	thousand Nm³/h	Nm³/GJ#	tonne/d
Steam consumption	0.379	10.304	14.642	198.9
Oxygen consumption**	0.611	9.368	13.311	321.2
Flux consumption	0.039	—	—	20.4

\# - Clean gas basis ie excluding ammonia, hydrogen chloride, hydrogen cyanide, and sulphur compounds.
- Gas volume is measured at 0°C and 1013·25 mbar (22.4 Nm³/kg mol)
** - Excluding all oxygen in air supplied to slag tap burners

Prepared by : J.M.F. 2nd June, 1989 Revision 0

TABLE 1 : Material Balance kg/h

INPUT	Total	C	H	O	N	S	Cl	Calcined mineral matter	Dust
Coal,dry ash free	21888	17835	1318	1210	363	1088	74	-	-
moisture	1798	-	201	1597	-	-	-	-	-
ash	3111	-	-	143	-	-	-	2968	-
Flux	848	91	-	242	-	-	-	515	-
Tar to the top of the bed	938	823	61	35	13	7	-	-	-
Dust to the top of the bed	195	-	-	-	-	-	-	-	195
Tar to tuyeres	1019	894	66	38	14	7	1	-	-
Oil to tuyeres	724	626	56	27	4	10	-	-	-
Naphtha to tuyeres	331	291	30	8	1	1	-	-	-
Oxygen to tuyeres	13999	-	-	13382	617	-	-	-	-
Steam to tuyeres	8287	-	927	7360	-	-	-	-	-
Burner gas	1288	74	8	348	859	-	-	-	-
TOTAL	54427	20632	2667	24389	1970	1113	76	3483	195

British Gas plc Midlands Research Station	PROJECT ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PRINCE COAL.	TITLE SLAGGING GASIFIER MATERIAL BALANCE	PREPARED BY J.M.F. DATE 2nd June. 1989	SHEET 1 of 6 REVISION 0

TABLE 1 Continued : Material Balance kg/h

OUTPUT	Total	C	H	O	N	S	Cl	Calcined mineral matter	Dust
Dry Gas (see sheet 3)	44292	17943	2074	21445	1740	1088	-	-	-
Steam	3187	-	357	2830	-	-	-	-	-
Tar	1957	1717	127	72	26	14	1	-	-
Oil	724	626	56	27	4	10	-	-	-
Naphtha	331	291	30	8	1	1	-	-	-
Phenols	32	25	2	5	-	-	-	-	-
Fatty Acids	5	3	-	1	-	-	-	-	-
NH3	109	-	19	-	89	-	-	-	-
HCl	77	-	2	-	-	-	74	-	-
HCN	19	8	1	-	10	-	-	-	-
Dust	195	-	-	-	-	-	-	-	195
Slag	3483	-	-	-	-	-	-	3483	-
C in Slag	19	19	-	-	-	-	-	-	-
TOTAL	54427	20632	2667	24389	1870	1113	76	3483	195

British Gas plc	PROJECT	TITLE	PREPARED BY J.M.F.	SHEET 2 of 6
Midlands Research Station	ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PRINCE COAL.	SLAGGING GASIFIER MATERIAL BALANCE	DATE 2nd June. 1989	REVISION 0

TABLE 1 Continued : Material Balance kg/h

OUTPUT : Dry Gas	Mol %	Total	C	H	O	N	S	Cl
H2	28.66	1268	-	1268	-	-	-	-
CH4	7.19	2531	1895	636	-	-	-	-
CO	57.00	35043	15026	-	20016	-	-	-
CO2	1.99	1925	525	-	1400	-	-	-
N2	2.83	1740	-	-	-	1740	-	-
C2H6	0.44	292	233	59	-	-	-	-
C2H4	0.16	97	83	14	-	-	-	-
C3 & C4	0.18	191	159	32	-	-	-	-
H2S	1.46	1094	-	65	-	-	1029	-
COS	0.08	110	22	-	29	-	59	-
CS2	0.00	1	-	-	-	-	-	-
TOTAL	100.00	44292	17943	2074	21445	1740	1088	-

British Gas plc Midlands Research Station	PROJECT ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PRINCE COAL.	TITLE SLAGGING GASIFIER MATERIAL BALANCE	PREPARED BY J.M.F. DATE 2nd June. 1989	SHEET 3 of 6 REVISION 0

TABLE 1 Continued : Heat Balance Megawatts

INPUT	Total Heat datum : 25°C	Potential Heat	Latent and Sensible Heat datum : 25°C	Temperature °C
Coal	212.82	212.82	-	25
Tar to the top of the bed	9.93	9.91	0.02	70
Dust to the top of the bed	1.78	1.78	-	70
Tar to tuyeres	10.78	10.76	0.02	70
Oil to tuyeres	7.89	7.87	0.02	70
Naphtha to tuyeres	4.13	4.12	0.01	70
Oxygen to tuyeres	0.40	-	0.40	135
Steam to tuyeres	7.16	-	7.16	400
Burner gas	0.80	0.80	-	25
TOTAL	255.69	248.07	7.62	-

British Gas plc Midlands Research Station	PROJECT ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PRINCE COAL.	TITLE SLAGGING GASIFIER HEAT BALANCE	PREPARED BY J.M.F. DATE 2nd June. 1989	SHEET 4 of 6 REVISION 0

TABLE 1 Continued : Heat Balance Megawatts

OUTPUT	Total Heat datum : 25°C	Potential Heat	Latent and Sensible Heat datum : 25°C	Temperature °C
Dry Gas (see sheet 6)	209.39	200.78	8.61	465
Steam	2.93	-	2.93	465
Tar	21.24	20.67	0.57	465
Oil	8.12	7.87	0.25	465
Naphtha	4.25	4.12	0.13	465
Phenols	0.30	0.29	0.01	465
Fatty Acids	0.04	0.04	-	465
NH3	0.71	0.68	0.03	465
HCl	0.01	-	0.01	465
HCN	0.08	0.07	-	465
Dust	1.81	1.78	0.03	465
Slag	0.87	0.87	0.01	30
Heat Losses	5.72	-	5.72	-
Heat of flux calcination	0.21	0.21	-	-
TOTAL	255.69	237.39	18.30	-

British Gas plc	PROJECT	TITLE	PREPARED BY	SHEET
Midlands Research Station	ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PRINCE COAL.	SLAGGING GASIFIER HEAT BALANCE	J.M.F.	5 of 6
			DATE 2nd June, 1989	REVISION 0

DATA SUMMARY SHEET
ATLANTIC COAL INSTITUTE ENQUIRY.

British Gas plc
Midlands Research Station

Coal Type : DONKIN COAL.

Coal Analysis

	Ultimate dry ash free wt %		Proximate as received wt %
C	82.48	Fixed Carbon	52.36
H	5.45	Volatile Matter	33.66
O	6.40	Moisture	6.50
N	1.41	Ash	7.48
S	4.12		
Cl	0.16		
	100.00		100.00

Coal feedrate	(as received)	611 tonne/d
Coal feedrate	(dry ash free)	21888 kg/h
Gasifier loading	(dry ash free)	4687 kg/h m²
No of gasifiers	(excludes spares)	1 x 2.44 m diameter
Gasification pressure		32.0 bar
Steam feed pressure		38.9 bar
Oxygen feed pressure		37.6 bar

Gasifier Performance

		dry gas ex gasifier
Gas output, million MJ/d	#	16.666
Gas output, million Nm³/d	#	1.142
Gas HHV, MJ/Nm³	#	14.59
Outlet temperature, °C		534

	kg/kg coal (dry ash free)	thousand Nm³/h	Nm³/GJ#	tonne/d
Steam consumption	0.382	10.384	14.954	200.4
Oxygen consumption**	0.616	9.440	13.594	323.7
Flux consumption	0.004	—	—	2.0

\# - Clean gas basis is excluding ammonia, hydrogen chloride, hydrogen cyanide, and sulphur compounds.
- Gas volume is measured at 0°C and 1013·25 mbar (22.4 Nm³/kg mol)
** - Excluding all oxygen in air supplied to slag tap burners

Prepared by : J.M.F. 2nd June, 1989 Revision 1

TABLE 1 Continued : Material Balance kg/h

OUTPUT : Dry Gas	Mol %	Total	C	H	O	N	S	Cl
H2	27.02	1173	-	1173	-	-	-	-
CH4	7.26	2509	1878	631	-	-	-	-
CO	58.99	35576	15255	-	20321	-	-	-
CO2	1.80	1702	465	-	1238	-	-	-
N2	2.82	1701	-	-	-	1701	-	-
C2H6	0.45	292	233	59	-	-	-	-
C2H4	0.16	97	82	14	-	-	-	-
C3 & C4	0.18	191	159	32	-	-	-	-
H2S	1.23	906	-	54	-	-	852	-
COS	0.07	92	19	-	24	-	49	-
CS2	0.00	-	-	-	-	-	-	-
TOTAL	100.00	44239	18091	1963	21583	1701	901	-

British Gas plc Midlands Research Station	PROJECT ACI ENQUIRY. ONE 8.0 FT.GASIFIER. DONKIN COAL.	TITLE SLAGGING GASIFIER MATERIAL BALANCE	PREPARED BY J.M.F. DATE 2nd June. 1989	SHEET 3 of 6 REVISION 0

TABLE 1 Continued : Heat Balance Megawatts

INPUT	Total Heat datum : 25°C	Potential Heat	Latent and Sensible Heat datum : 25°C	Temperature °C
Coal	209.05	209.05	-	25
Tar to the top of the bed	9.93	9.91	0.02	70
Dust to the top of the bed	1.78	1.78	-	70
Tar to tuyeres	9.12	9.11	0.02	70
Oil to tuyeres	7.25	7.24	0.01	70
Naphtha to tuyeres	4.13	4.12	0.01	70
Oxygen to tuyeres	0.40	-	0.40	135
Steam to tuyeres	7.22	-	7.22	400
Burner gas	0.80	0.80	-	25
TOTAL	249.68	242.00	7.68	-

TABLE 1 Continued : Heat Balance Megawatts

OUTPUT	Total Heat datum : 25°C	Potential Heat	Latent and Sensible Heat datum : 25°C	Temperature °C
Dry Gas (see sheet 6)	207.13	197.28	9.85	534
Steam	2.86	-	2.86	534
Tar	19.61	19.02	0.59	534
Oil	7.51	7.24	0.27	534
Naphtha	4.27	4.12	0.15	534
Phenols	0.30	0.29	0.01	534
Fatty Acids	0.05	0.04	-	534
NH_3	0.63	0.60	0.03	534
HCl	-	-	-	534
HCN	0.08	0.07	-	534
Dust	1.81	1.78	0.04	534
Slag	0.53	0.53	-	30
Heat Losses	4.87	-	4.87	-
Heat of flux calcination	0.02	0.02	-	-
TOTAL	249.68	230.99	18.69	-

British Gas plc Midlands Research Station	PROJECT ACI ENQUIRY. ONE 8.0 FT.GASIFIER. DONKIN COAL.	TITLE SLAGGING GASIFIER HEAT BALANCE	PREPARED BY J.M.F. DATE 2nd June, 1989	SHEET 5 of 6 REVISION 0

BRITISH GAS plc

Performance Predictions of the BGL gasifier for Atlantic Coal Institute

Appendix 3

Performance prediction of the BGL Gasifier on Phalen Harbour coal.

Notes

1. Gasifier Loading = 960 lb (DAF)/sq ft h

2. Cold Clean Gas Efficiency = 92.89%

3. The IGT correlation was used to calculate the coal HHV since the quoted CV gave an unrealistically high offtake temperature for the gasifier.

4. Desulphurised product gas was assumed for the burner gas with air as the oxidant.

DATA SUMMARY SHEET
ATLANTIC COAL INSTITUTE ENQUIRY.

British Gas plc
Midlands Research Station

Coal Type : PHALEN COAL.

Coal Analysis

	Ultimate dry ash free wt %		Proximate as received wt %
C	85.14	Fixed Carbon	54.16
H	5.47	Volatile Matter	32.67
O	5.36	Moisture	8.50
N	1.58	Ash	4.67
S	2.21		
Cl	0.24		
	100.00		100.00

Coal feedrate	(as received)	605 tonne/d
Coal feedrate	(dry ash free)	21888 kg/h
Gasifier loading	(dry ash free)	4687 kg/h m²
No of gasifiers	(excludes spares)	1 x 2.44 m diameter
Gasification pressure		32.0 bar
Steam feed pressure		38.9 bar
Oxygen feed pressure		37.6 bar

Gasifier Performance

		dry gas ex gasifier
Gas output, million MJ/d	#	17.323
Gas output, million Nm³/d	#	1.192
Gas HHV, MJ/Nm³	#	14.53
Outlet temperature, °C		587

	kg/kg coal (dry ash free)	thousand Nm³/h	Nm³/GJ#	tonne/d
Steam consumption	0.398	10.821	14.992	208.9
Oxygen consumption**	0.642	9.937	13.629	337.3
Flux consumption	0.007	—	—	3.8

\# - Clean gas basis ie excluding ammonia, hydrogen chloride, hydrogen cyanide, and sulphur compounds.
- Gas volume is measured at 0°C and 1013·25 mbar (22.4 Nm³/kg mol)
** - Excluding all oxygen in air supplied to slag tap burners

TABLE 1 : Material Balance kg/h

INPUT	Total	C	H	O	N	S	Cl	Calcined mineral matter	Dust
Coal,dry ash free	21888	18636	1197	1173	346	484	53	-	-
moisture	2143	-	240	1903	-	-	-	-	-
ash	1177	-	-	55	-	-	-	1122	-
Flux	160	17	-	46	-	-	-	97	-
Tar to the top of the bed	938	823	61	35	13	7	-	-	-
Dust to the top of the bed	195	-	-	-	-	-	-	-	195
Tar to tuyeres	684	600	44	25	9	5	-	-	-
Oil to tuyeres	600	518	47	23	3	9	-	-	-
Naphtha to tuyeres	331	291	30	9	1	1	-	-	-
Oxygen to tuyeres	14701	-	-	14053	648	-	-	-	-
Steam to tuyeres	8703	-	974	7729	-	-	-	-	-
Burner gas	1291	75	8	349	859	-	-	-	-
TOTAL	52810	20960	2609	25397	1879	505	54	1220	195

British Gas plc Midlands Research Station	PROJECT ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PHALEN COAL.	TITLE SLAGGING GASIFIER MATERIAL BALANCE	PREPARED BY J.M.F. DATE 2nd June, 1989	SHEET 1 of 6 REVISION 0

TABLE 1 Continued : Material Balance kg/h

OUTPUT	Total	C	H	O	N	S	Cl	Calcined mineral matter	Dust
Dry Gas (see sheet 3)	45206	18685	2014	22261	1762	484	-	-	-
Steam	3422	-	382	3039	-	-	-	-	-
Tar	1622	1428	105	60	22	12	1	-	-
Oil	600	518	47	23	3	9	-	-	-
Naphtha	331	291	30	8	1	1	-	-	-
Phenols	32	25	2	5	-	-	-	-	-
Fatty Acids	5	3	-	1	-	-	-	-	-
NH_3	99	-	18	-	81	-	-	-	-
HCl	54	-	1	-	-	-	53	-	-
HCN	19	9	1	-	10	-	-	-	-
Dust	195	-	-	-	-	-	-	-	195
Slag	1220	-	-	-	-	-	-	1220	-
C in Slag	-	7	-	-	-	-	-	-	-
TOTAL	52810	20960	2600	25397	1979	505	54	1220	195

British Gas plc Midlands Research Station	PROJECT ACI ENQUIRY. ONE 9.0 FT.GASIFIER. PHALEN COAL.	TITLE SLAGGING GASIFIER MATERIAL BALANCE	PREPARED BY J.M.F. DATE 2nd June. 1989	SHEET 2 of 6 REVISION 0

TABLE 1 Continued : Material Balance kg/h

OUTPUT : Dry Gas	Mol %	Total	C	H	O	N	S	Cl
H2	27.87	1254	-	1254	-	-	-	-
CH4	6.96	2493	1866	626	-	-	-	-
CO	59.54	37234	15966	-	21268	-	-	-
CO2	1.37	1347	368	-	979	-	-	-
N2	2.82	1762	-	-	-	1762	-	-
C2H6	0.43	292	233	59	-	-	-	-
C2H4	0.16	97	83	14	-	-	-	-
C3 & C4	0.18	191	159	32	-	-	-	-
H2S	0.64	484	-	29	-	-	456	-
COS	0.04	52	10	-	14	-	28	-
CS2	0.00	-	-	-	-	-	-	-
TOTAL	100.00	45206	18695	2014	22261	1762	484	-

British Gas plc	PROJECT	TITLE	PREPARED BY	SHEET
Midlands Research Station	ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PHALEN COAL.	SLAGGING GASIFIER MATERIAL BALANCE	J.M.F.	3 of 6
			DATE 2nd June. 1989	REVISION 0

TABLE 1 Continued : Heat Balance Megawatts

OUTPUT	Total Heat datum : 25°C	Potential Heat	Latent and Sensible Heat datum : 25°C	Temperature °C
Dry Gas (see sheet 6)	214.13	202.85	11.28	587
Steam	3.39	-	3.39	587
Tar	17.71	17.13	0.58	587
Oil	6.79	6.52	0.27	587
Naphtha	4.29	4.12	0.17	587
Phenols	0.31	0.29	0.01	587
Fatty Acids	0.04	0.04	-	587
NH3	0.66	0.62	0.04	587
HCl	0.01	-	0.01	587
HCN	0.08	0.07	-	587
Dust	1.82	1.78	0.04	587
Slag	0.32	0.32	-	30
Heat Losses	4.59	-	4.59	-
Heat of flux calcination	0.04	0.04	-	-
TOTAL	254.19	233.79	20.40	-

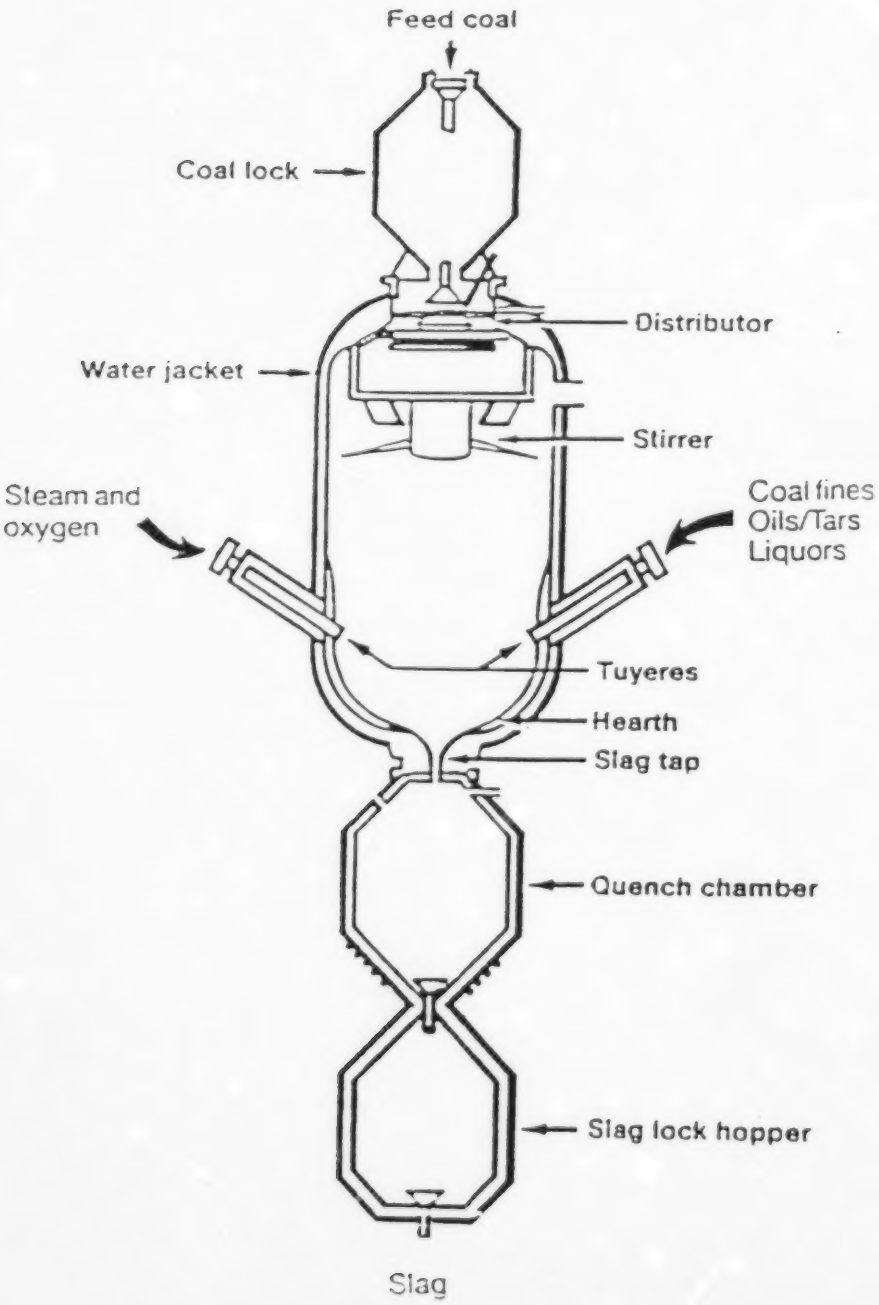
British Gas plc Midlands Research Station	PROJECT ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PHALEN COAL.	TITLE SLAGGING GASIFIER HEAT BALANCE	PREPARED BY J.M.F. DATE 2nd June, 1989	SHEET 5 of 6 REVISION 0

TABLE 1 Continued : Heat Balance Megawatts

OUTPUT : Dry Gas	Total Heat datum : 25°C	Potential Heat	Latent and Sensible Heat datum : 25°C	Temperature °C
H2	52.24	49.39	2.85	587
CH4	39.69	38.44	1.25	587
CO	110.82	104.49	6.33	587
CO2	0.22	-	0.22	587
N2	0.30	-	0.30	587
C2H6	4.33	4.20	0.12	587
C2H4	1.40	1.36	0.04	587
C3 & C4	2.70	2.62	0.08	587
H2S	2.31	2.22	0.09	587
COS	0.14	0.13	0.01	587
CS2	-	-	-	587
TOTAL	214.13	202.85	11.28	-

British Gas plc	PROJECT	TITLE	PREPARED BY	SHEET
Midlands Research Station	ACI ENQUIRY. ONE 8.0 FT.GASIFIER. PHALEN COAL.	SLAGGING GASIFIER HEAT BALANCE	J.M.F.	6 of 6
			DATE 2nd June. 1989	REVISION 0

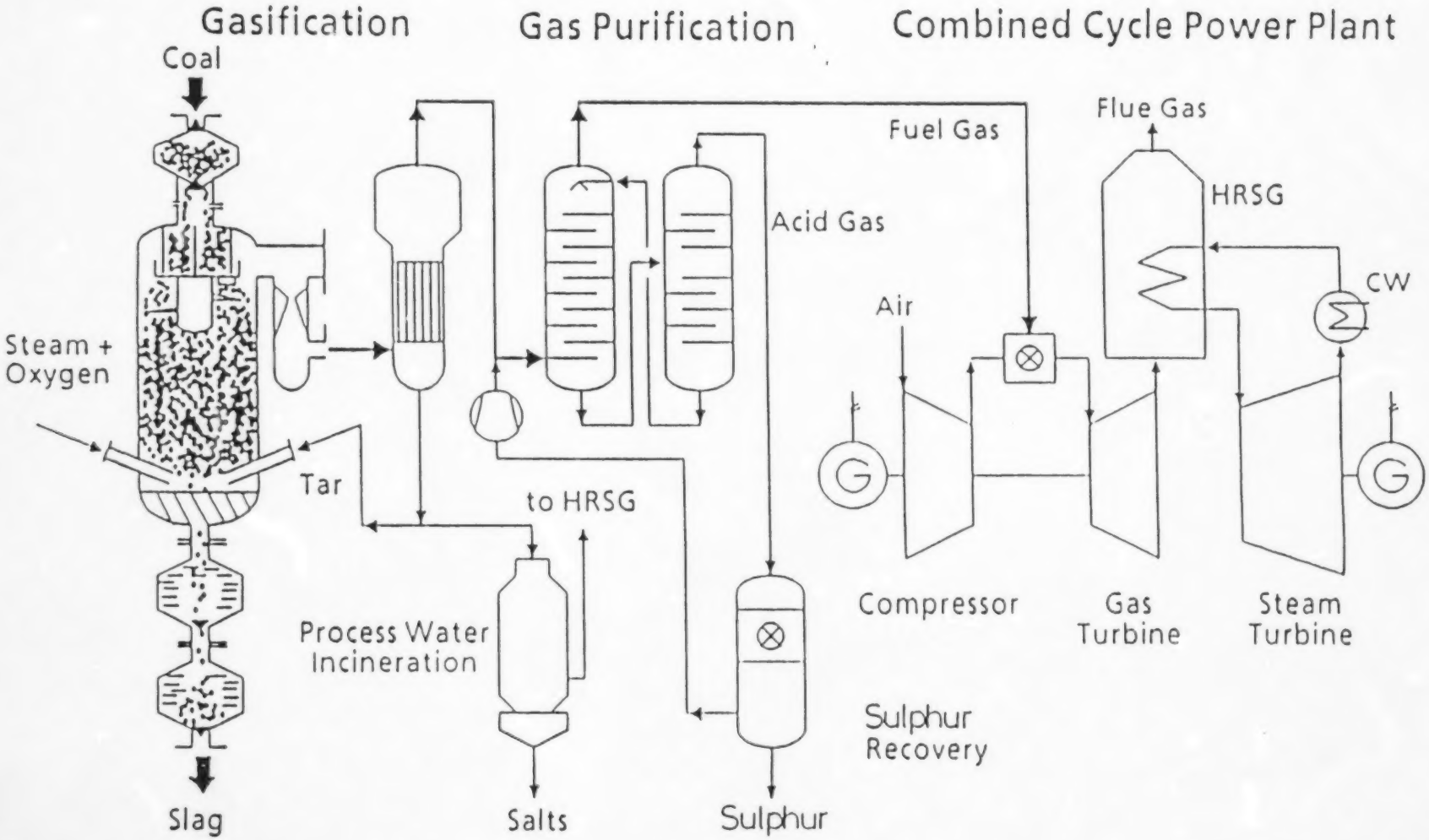
Figure 1 The British Gas Lurgi Gasifier



BGL - Gasification Combined Cycle

Figure 2

Base Scheme



KELLOGG

The M. W. Kellogg Company

Three Greenway Plaza
Houston, Texas 77046-0395

(713) 960-

21-APR-89

Mr. Donald A. Anderson
Atlantic Coal Institute
Box 1594
Sydney, Nova Scotia
Canada B1P 6R8

Re: MWK Project 6539
IGCC Study
Progess Report for the week ending 21 April 1989

Dear Mr. Anderson:

We have completed computer simulations of KRW based IGCC power plants for the three coals specified by you. Documentation of the three cases is being prepared, and will be shipped to you early next week.

The results for all three coals are similar and are summarized below:

	Prince Hub	Donkin Harbour	Phelan Harbour
Coal Rate, ST/D	2005	1988	1793
Net Power, MWe	254.2	257.1	250.3
Net Plant Heat Rate, Btu(LHV)/kWh	8129	8206	8065
No. of Gasifier Trains	3	3	3

We hope the above will permit you to begin your analysis. Overall material balances and Block Flow Diagrams will be mailed by 26 April.

If you have any questions, please do not hesitate to call.

Regards,

I.N. Banchik
Process Manager

KELLOGG

The M. W. Kellogg Company

Three Greenway Plaza
Houston, Texas 77046-0395

(713) 960-

27 April 1989

Mr. Donald A. Anderson
Atlantic Coal Institute
Box 1594
Sydney, Nova Scotia
Canada B1P 6R8

Re: MWK Project 6539
IGCC Study
Final Report

Dear Mr. Anderson:

In accordance with your Purchase Order Number 63960 and your letter of approval dated March 22, 1989, we are pleased to enclose the following documents:

1. Block Flow Diagram
2. KRW IGCC Performance Summary
3. Overall Material Balance - Prince Hub Coal
4. Overall Material Balance - Donkin Harbour Coal
5. Overall Material Balance - Phelan Harbour Coal

These documents represent the results of three computer simulations of KRW based IGCC facilities. In each case, the capacity of the gasifier island was set to fully load a GE MS7001F advanced gas turbine at a 32°F ambient temperature. To satisfy this capacity requirement, the gasifier island in all three cases will contain three KRW gasification trains with hot gas cleanup.

Coal suitability for gasification in a KRW gasifier can only be determined by bench scale and pilot plant testing. Bench scale testing would provide a measure of the coal's reactivity and ash fusion characteristics under gasifier conditions. Pilot plant tests would permit determination of optimum gasifier operating conditions for the coal in question. The bases for these studies were the coal analyses provided by you. Gasifier performance parameters were taken from pilot plant operation on other coals which were similar in analysis.

Other bases which should be noted are as follows:

1. Unconverted carbon leaving the gasifier is almost totally combusted in a sulfator. This device, which is essentially a sulfide roaster, is used to oxidize the calcium sulfide in the spent gasifier solids to inert calcium sulfate. Concurrent combustion of the unconverted carbon results in the production of additional high pressure steam for power generation.

2. The sulfur content of the fuel gas leaving the gasifier island is less than 20 ppm. This corresponds to a desulfurization efficiency of greater than 99% in all these cases.

3. For the purpose of this study, it has been assumed that the coal will be delivered sized, 1/4" x 0, and surface dry, 5% moisture. Hence, the energy necessary to crush and dry the raw coal has not been accounted for in the auxiliary power requirement. It is recommended that, when designing the crusher/dryer, sulfator exhaust heat be utilized to provide the energy for drying the coal.

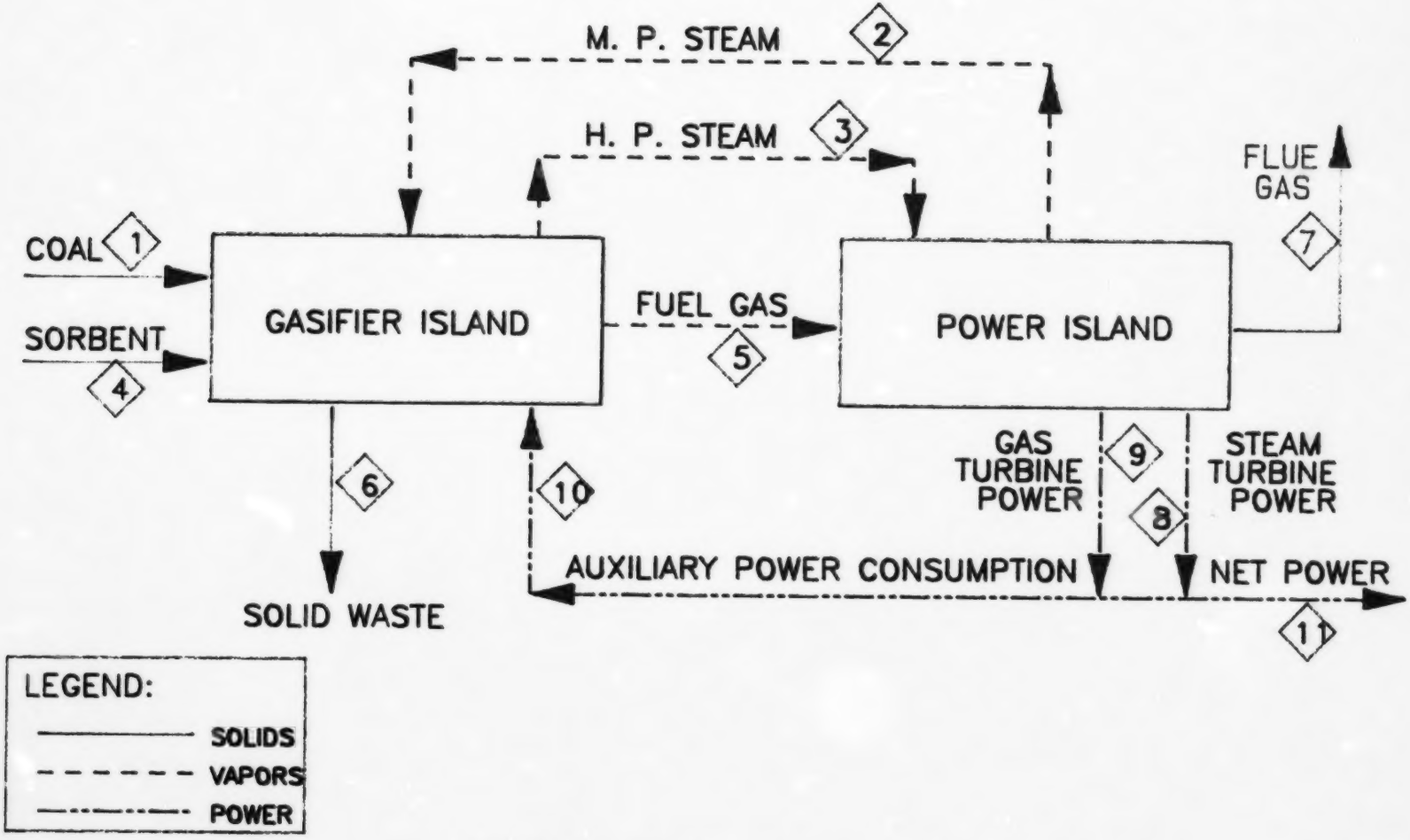
4. The sorbent was assumed to be a magnesian limestone containing approximately 65% calcium carbonate and 20% magnesium carbonate.

We hope this information satisfies your requirements. If we may be of any further service, please do not hesitate to contact us.

Very truly yours,

I. N. Banchik
Process Manager

INB/ss



BLOCK FLOW DIAGRAM
ATLANTIC COAL INSTITUTE

KRW IGCC PERFORMANCE SUMMARY

PERFORMANCE PARAMETERS	PRINCE HUB	DONKIN HARBOUR	PHELAN HARBOUR
Coal Rate (as fed), Lb/Hr	167,119	165,698	149,390
Limestone Rate, Lb/Hr	61,418	58,552	25,730
Solid Waste, Lb/Hr	68,382	63,449	30,493
POWER GENERATED			
Gas Turbine, kW (STREAM 9)	182,728	183,718	181,226
Steam Turbine, kW (STREAM 8)	82,321	85,148	79,812
Total Power, kW	265,049	268,866	261,038
Power Consumed, kW (STREAM 10)	12,290	13,085	11,327
NET POWER EXPORT, kW	252,759	255,781	249,711

OVERALL MATERIAL BALANCE

STREAM NUMBER IDENTIFICATION	Molecular Weight	1 Prepared Coal Feedstock		2 M.P. Steam to Gasif. Section		3 H.P.Steam Export From Gasifier		4 Sorbent to Gasifier		5 Fuel Gas to Gas Turbine		6 Solid Waste		7 HRSG Flue Gas	
		Wt%	Lb/Hr	Vol%	Lb/Hr	Vol%	Lb/Hr	Wt%	Lb/Hr	Vol%	Lb/Hr	Wt%	Lb/Hr	Vol%	Lb/Hr
GASES:															
Carbon Monoxide	28.010									8.84	101,584			0.00	0
Hydrogen	2.016									20.64	17,077			0.00	0
Carbon Dioxide	44.010									13.48	243,384			7.56	426,995
Methane	16.043									1.28	8,430			0.00	0
Nitrogen	28.013									34.92	401,424			67.91	2,441,162
Argon	39.948									0.00	0			0.68	34,464
Oxygen	31.999									0.00	0			9.69	397,972
Ammonia	17.031									0.01	101			0.00	0
Hydrogen Sulfide	34.080									0.00	28			0.00	0
Carbonyl Sulfide	60.075									0.00	0			0.00	0
Sulfur Dioxide	64.063									0.00	0			0.00	0
Water Vapor	18.015			100.00	212,522	100.00	466,010			20.02	153,897			14.16	327,230
TOTAL GASES				100.00	212,522	100.00	466,010			100.00	925,925			100.00	3,627,823
Gas Flow, Lb Moles/Hr					11,797		25,868				41,032				128,309
Molecular Weight					18.015		18.015				22.566				28.274
Gas Volume, ACFM					6,455		1,979				30,610				1,256,870
Gas Volume, SCFM											259,575				811,405
LIQUIDS:															
Water	18.015														
SOLIDS:															
Carbon	12.011	68.00	113,641					10.70	6,573			0.10	78		
Hydrogen	1.008	5.03	8,406					0.00	0			0.00	0		
Oxygen	16.000	4.62	7,721					42.70	26,226			0.00	0		
Nitrogen	14.008	1.38	2,306					0.00	0			0.00	0		
Sulfur	32.064	4.14	6,919					0.00	0			9.20	6,898		
Ash/Inerts		11.83	19,770					46.60	28,619			90.69	67,985		
Moisture	18.015	5.00	8,356					0.00	0			0.00	0		
TOTAL SOLIDS		100.00	167,119					100.00	61,418			100.00	74,961		
TOTAL FLOW, Lb/Hr			167,119		212,522		466,010		61,418		925,925		74,961		3,627,823
Pressure, psia			14.6		450.0		1600.0		14.6		350.0		14.6		14.6
Temperature, F			32.0		950.0		605.0		32.0		1000.0		120.0		340.0

OVERALL MATERIAL BALANCE

STREAM NUMBER IDENTIFICATION	Molecular Weight	1 Prepared Coal Feedstock		2 M.P. Steam to Gasif. Section		3 H.P.Steam Export From Gasifier		4 Sorbent to Gasifier		5 Fuel Gas to Gas Turbine		6 Solid Waste		7 HRSG Flue Gas	
		Wt%	Lb/Hr	Vol%	Lb/Hr	Vol%	Lb/Hr	Wt%	Lb/Hr	Vol%	Lb/Hr	Wt%	Lb/Hr	Vol%	Lb/Hr
GASES:															
Carbon Monoxide	28.010									8.66	184,492			0.00	0
Hydrogen	2.016									19.39	16,834			0.00	0
Carbon Dioxide	44.010									13.78	261,173			7.56	449,298
Methane	16.043									1.23	8,560			0.00	0
Nitrogen	28.013									36.53	440,754			67.91	2,441,647
Argon	39.948									0.00	0			0.68	33,861
Oxygen	31.999									0.00	0			9.89	385,945
Ammonia	17.031									0.01	93			0.00	0
Hydrogen Sulfide	34.080									0.00	29			0.00	0
Carbonyl Sulfide	60.075									0.00	0			0.00	0
Sulfur Dioxide	64.063									0.00	0			0.00	0
Water Vapor	18.015			100.00	217,680	100.00	548,194			20.40	158,265			14.16	329,509
TOTAL GASES				100.00	217,680	100.00	548,194			100.00	990,140			100.00	3,640,260
Gas Flow, Lb Moles/Hr					12,083		30,436				43,070				128,586
Molecular Weight					18.015		18.015				22.989				28.310
Gas Volume, ACFM					6,612		2,328				32,130				1,256,870
Gas Volume, SCFM											272,466				813,156
LIQUIDS:															
Water	18.015														
SOLIDS:															
Carbon	12.011	72.20	119,636					10.69	6,262			0.11	69		
Hydrogen	1.008	4.75	7,870					0.00	0			0.00	0		
Oxygen	16.000	5.61	9,297					42.74	25,023			0.00	0		
Nitrogen	14.008	1.23	2,042					0.00	0			0.00	0		
Sulfur	32.064	3.60	5,967					0.00	0			9.36	5,936		
Ash/Inerts		7.60	12,593					46.57	27,267			90.54	57,444		
Moisture	18.015	5.00	8,285					0.00	0			0.00	0		
TOTAL SOLIDS		100.00	165,698					100.00	58,552			100.00	63,449		
TOTAL FLOW, Lb/Hr			165,698		217,680		548,194		58,552		990,140		63,449		3,640,260
Pressure, psia			14.6		450.0		1600.0		14.6		350.0		14.6		14.6
Temperature, F			32.0		950.0		605.0		32.0		1000.0		120.0		350.0

OVERALL MATERIAL BALANCE

STREAM NUMBER / IDENTIFICATION	Molecular Weight	1 Prepared Coal Feedstock		2 M.P. Steam to Gasif. Section		3 H.P.Steam Export From Gasifier		4 Sorbent to Gasifier		5 Fuel Gas to Gas Turbine		6 Solid Waste		7 HRSG Flue Gas	
		Wt%	Lb/Hr	Vol%	Lb/Hr	Vol%	Lb/Hr	Wt%	Lb/Hr	Vol%	Lb/Hr	Wt%	Lb/Hr	Vol%	Lb/Hr
GASES:															
Carbon Monoxide	28.010									9.04	102,796			0.00	0
Hydrogen	2.016									20.99	17,192			0.00	0
Carbon Dioxide	44.010									13.18	235,635			7.56	418,910
Methane	16.043									1.17	7,600			0.00	0
Nitrogen	28.013									35.34	402,071			67.91	2,440,798
Argon	39.948									0.66	0			0.68	34,698
Oxygen	31.999									0.00	0			9.69	399,390
Ammonia	17.031									0.01	101			0.00	0
Hydrogen Sulfide	34.080									0.00	28			0.00	0
Carbonyl Sulfide	60.075									0.00	0			0.00	0
Sulfur Dioxide	64.063									0.00	0			0.00	0
Water Vapor	18.015			100.00	215,872	100.00	465,928			20.27	148,318			14.16	320,958
TOTAL GASES				100.00	215,872	100.00	465,928			100.00	913,741			100.00	3,614,754
Gas Flow, Lb Moles/Hr					11,983		25,863				40,618				127,352
Molecular Weight					18.015		18.015				22.496				28.384
Gas Volume, ACFM					6,557		1,979				30,301				1,259,816
Gas Volume, SCFM											256,955				865,353
LIQUIDS:															
Water	18.015														
SOLIDS:															
Carbon	12.011	76.94	114,946					10.70	2,752			0.24	73		
Hydrogen	1.008	5.50	8,217					0.00	0			0.00	0		
Oxygen	16.000	9.29	13,878					42.74	10,996			0.00	0		
Nitrogen	14.008	1.43	2,129					0.00	0			0.00	0		
Sulfur	32.064	1.99	2,977					0.00	0			9.67	2,950		
Ash/Inerts		4.85	7,243					46.57	11,982			90.09	27,470		
Moisture	18.015	5.00	7,470					0.00	0			0.00	0		
TOTAL SOLIDS		105.00	149,390					100.00	25,730			100.00	30,493		
TOTAL FLOW, Lb/Hr			149,390		215,872		465,928		25,730		913,741		30,493		3,614,754
Pressure, psia			14.6		450.0		1600.0		14.6		350.0		14.6		14.6
Temperature, F			32.0		950.0		605.0		32.0		1000.0		120.0		350.0

Post Office Box 334
Madison, Pennsylvania 15663-0334
(412) 722-6001
TWX 510-468-6572

KRW Energy Systems Inc.

May 17, 1989
MBB-89-065

Dr. Dieter Birk, Director
Atlantic Coal Institute
Box 1594
Sydney, Nova Scotia
Canada B1P 6R8

Dear Dr. Birk:

This letter pertains to my discussion with Mr. Glynn Campbell on May 9, 1989 regarding results of the recent KRW technology-based IGCC study based on three Nova Scotia coals; the study was conducted by KRW's parent firm, The M. W. Kellogg Company. Shortly you will receive the additional information requested by Mr. Campbell. This letter also proposes to add another case to the study.

The performance of the IGCC power plants based on KRW gasification/hot gas cleanup technology is quite impressive with net efficiencies being approximately 41 percent (HHV basis); the scope of these studies did not permit system optimization. The KRW technology in these power plants uses: (1) a limestone feed to the coal gasifier to accomplish about 90-94 percent desulfurization within the gasifier; (2) hot particulate removal; and (3) desulfurization of the fuel gas to about 5-10 ppm by means of regenerable zinc ferrite. The product is extremely clean gas turbine fuel that results in very low atmospheric emissions of sulfur dioxide - in fact, much lower than necessary to meet Canada's environmental requirements.

KRW proposes to extend the study to include a case that would eliminate the zinc ferrite desulfurization subsystem. We expect that this change would result in a small improvement in net IGCC efficiency while still meeting the requirement of at least 90 percent sulfur removal.

Other benefits of eliminating the zinc ferrite system would be:

- Lower capital costs-due to elimination of some equipment and a size reduction of other equipment.

- Lower operating costs-due to the use of less limestone because (1) there would be no zinc ferrite regeneration gases routed to the gasifier for desulfurization by limestone, and (2) there would be less solid waste for disposal.

- Reduced water use because steam added to the fuel gas to prevent carbon deposition (catalyzed by zinc ferrite) would be eliminated.

If you authorize us to proceed, Kellogg will study the additional case involving elimination of zinc ferrite - with gasifier operation on Prince Hub coal, or another if preferred. We will provide the same inputs to you as for the current study, namely:

Coal requirement
Number of gasifier trains required
Plant effluent quantities and characteristics
Overall material balance
Suitability of coal for service

The net firm price to Atlantic Coal Institute for this additional case is U.S.$1,250. As previously, Kellogg will cost-share a portion of the total costs. We will begin the work as soon as your authorization is received.

Please contact me if there are any questions. My new telephone number is (412) 469-6148. The KRW Fax number is (412) 469-6178 and our new address is: P. O. Box 9886, Pittsburgh, PA 15227.

Very truly yours,

Morton B. Blinn

Morton B. Blinn
Manager Business Development

cc: W. M. Campbell

KELLOGG

The M. W. Kellogg Company

Three Greenway Plaza
Houston, Texas 77046-0395

(713) 960- 2606

22 May 1989

Dr. Dieter Birk, Director
Atlantic Coal Institute
Box 1594
Sydney, Nova Scotia
Canada B1P 6R8

Re: MWK Project 6539
IGCC Study
Additional Information

Dear Dr. Birk:

Further to our letter of 27 April 1989 and subsequent telephone conversations between your Mr. G. Campbell and our Mr. M. B. Blinn, we are pleased to provide you with the following information:

1. We wish to confirm the ability of the KRW gasification-desulfurization technology to process the three coals that were the basis for the IGCC studies recently completed; specifically Prince Hub, Donken Harbour, and Phelan Harbour coals. The KRW pilot plant has accumulated more than 7000 hours of successful operation on bituminous coals, approximately 3000 of which have been in the hot gas cleanup mode. The coals processed span the range of properties provided for the three Nova Scotia coals, and the design of the Appalachian Project, a 60 MW IGCC plant based on KRW technology, is based on a similar coal.

2. A revised Block Flow Diagram with streams 8, 9, 10 and 11 deleted as these values are not included in the overall material balance tables provided previously. The data is, however, presented in the KRW IGCC Performance Summary Table in your possession.

3. Net plant heat rates on an HHV basis are 8477, 8526 and 8352 Btu/KWH for Prince Hub, Donken Harbour and Phelan Harbour respectively. The HHV of each coal was calculated by the Dulong Formula. Please note that the designs produced in the subject study were not optimized. We expect that optimization of both the gasifier island and the power island could result in a significant reduction in the net plant heat rate, perhaps 300-500 Btu/KWH.

4. Gasifier air requirements for the three cases are 517740, 568814 and 518954 lb/hr for Prince Hub, Donken Harbour and Phelan Harbour coals respectively.

5. The current designs are predicated on 20 ppm by volume of HS in the fuel gas exiting the zinc ferrite desulfurizer. Based on this level of sulfur in the fuel gas, the HRSG stack gas will contain approximately 53 lb/hr of SO_2 (15 ppm by weight).

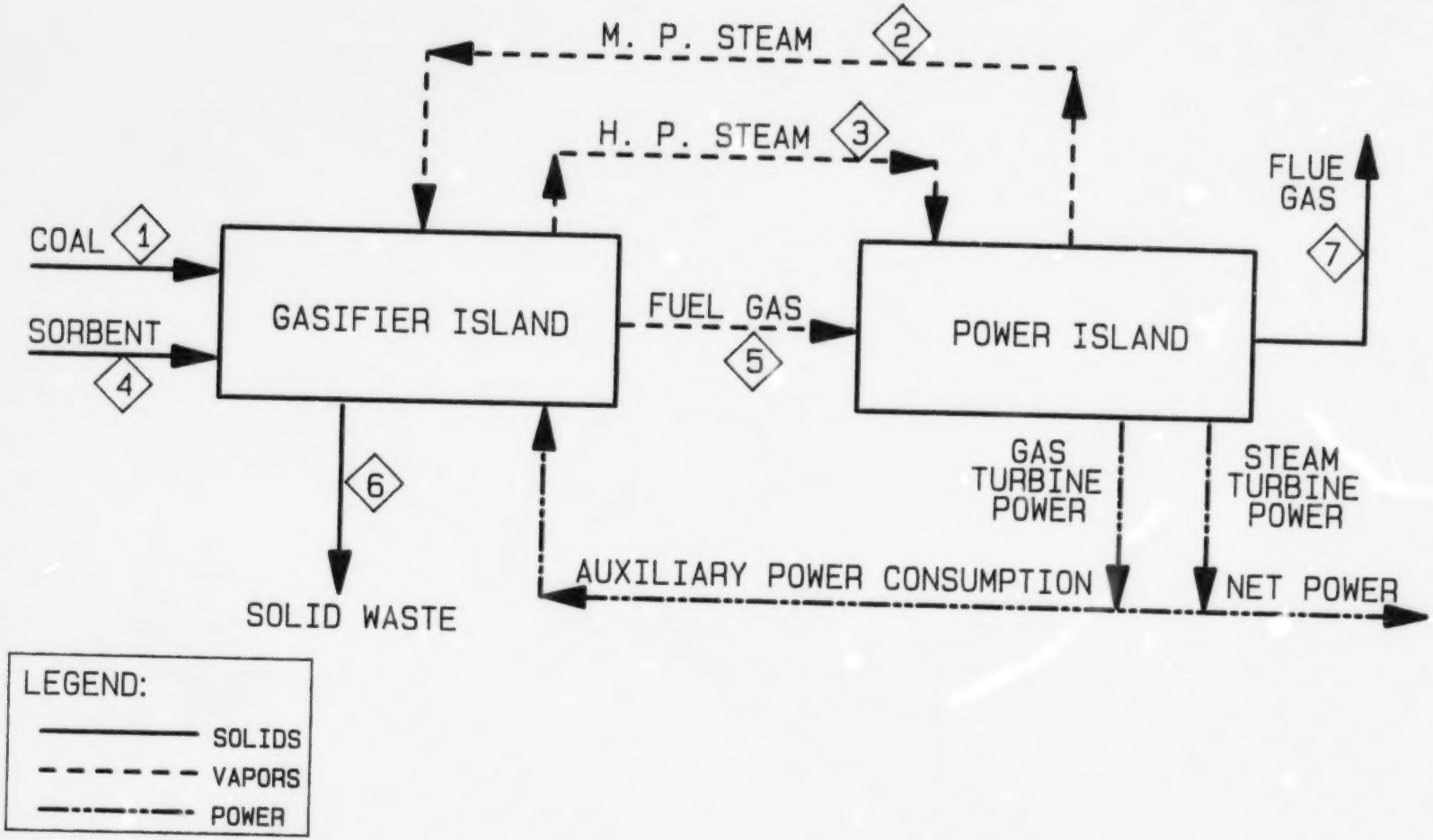
6. Atmospheric emissions of CO and NOx are a function of the design of the gas turbine. General Electric has informed us that with the fuel gas produced by the KRW gasifier they expect to meet all regulatory requirements with regard to CO and NOx. Additionally, with the MS7001F they believe that requisite levels of NOx emission can be achieved without the need for additional steam injection.

Should you require any additional information, please do not hesitate to contact us.

Very truly yours,

I. N. Banchik
Process Manager

INB/ss



BLOCK FLOW DIAGRAM
ATLANTIC COAL INSTITUTE

DOW CHEMICAL U.S.A.

400 WEST SAM HOUSTON PARKWAY SOUTH
P.O. BOX 3387
HOUSTON, TEXAS 77253-3387

April 5, 1989

Mr. Donald A. Anderson
Atlantic Coal Institute
P. O. Box 1594
Sydney, Nova Scotia, Canada

Dear Mr. Anderson:

The computer simulations requested by the Atlantic Coal Institute of the Dow-based integrated gasification combined cycle (IGCC) power plant have been completed for the three coals given in your letter of February 24, 1989. A diagram of the process modeled is given in Figure 1. Performance data for an IGCC power plant based on the Dow gasification process for the Prince Hub, Donkin-Morien Harbour, and Phalen-Lingan Harbour coals are given in Tables 1-3 respectively.

The three coals given for this study appear to be suitable for processing in the Dow gasifier. From the moisture content given for the as received coal, slurry concentrations of 64.66 weight percent solids would be expected. Process efficiency increases with increasing slurry concentration. A review of the ash compositions indicates sufficiently low T 250 temperatures, the temperature of which the ash has a viscosity of 250 poise. The ash T-250 temperature is an indicator of the degree of fluidity the ash would have at the gasification temperature used in the first stage. The molten ash produced must be fluid enough to drain into the water quench section at the bottom of the gasifier.

The parameters used for the gas turbine are based on data available for the GE MS7001F gas turbine. An exact comparison with other GE data may differ slightly. The increased performance of the advanced gas turbine design is included within the modeling.

Please call and advise my office if the information fulfills your requirements as per your letter of 2-24-89.

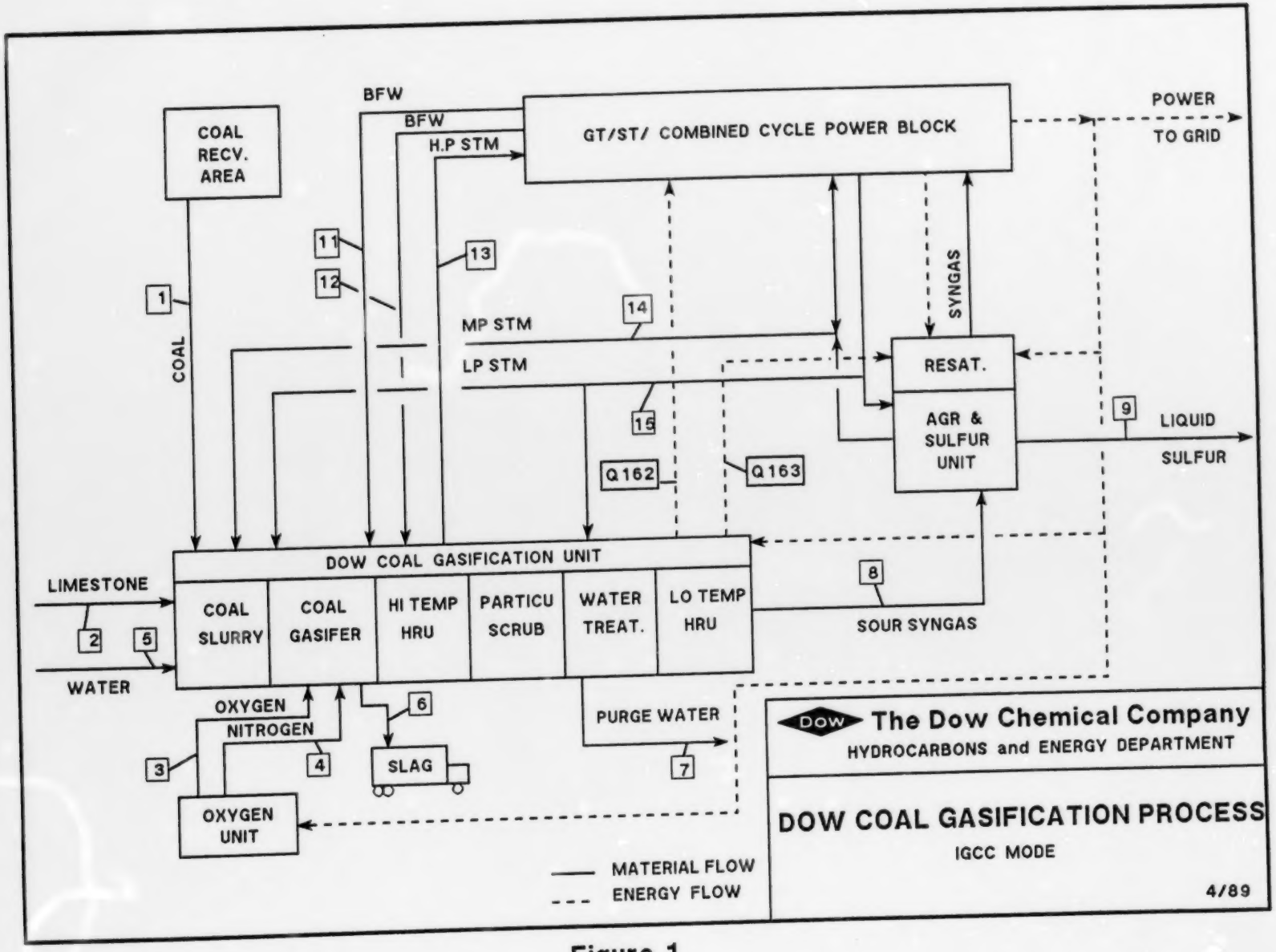
If you have any questions, please do not hesitate to call me. My phone number is (713) 978-2383.

Sincerely yours,

David L. Breton/pg

David L. Breton
Process Associate

psf
attachments


BFW
BFW
H.P STM
GT/ST/ COMBINED CYCLE POWER BLOCK
POWER
TO GRID
COAL
RECV.
AREA
11
12
13
1
COAL
MP STM
14
LP STM
SYNGAS
RESAT.
AGR &
SULFUR
UNIT
9
LIQUID
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Q162
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SLURRY
COAL
GASIFER
HI TEMP
HRU
PARTICU
SCRUB
WATER
TREAT.
LO TEMP
HRU
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SOUR SYNGAS
OXYGEN
NITROGEN
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4
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SLAG
PURGE WATER
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OXYGEN
UNIT
MATERIAL FLOW
ENERGY FLOW
DOW
The Dow Chemical Company
HYDROCARBONS and ENERGY DEPARTMENT
DOW COAL GASIFICATION PROCESS
IGCC MODE
4/89

Figure 1

TABLE 1
MATERIAL AND ENERGY BALANCE
ONE MODULE - ~200 MW GAS IGCC

Coal Source: Prince Mine / Hub Seam

MATERIAL

Process Stream		lb/Hr
1.	Coal	149800
	1.1 Dry Solids	139800
	Moisture	10000
2.	Limestone	None
3.	95% Oxygen	135400
	3.1 O_2	128800
	3.2 N_2 + A	6600
4.	Nitrogen	2000
5.	Water	68800
6.	Slag	24000
	6.1 Dry Solids	17800
	6.2 Moisture	6200
7.	Water Purge	47100
8.	Sour Syngas	288562
	8.1 H_2S	6130
	8.2 H_2	10140
	8.3 NH_3	2
	8.4 HCN	20
	8.5 CO	178520
	8.6 COS	470
	8.7 CH_4	780
	8.8 CO_2	81040
	8.9 N_2 + A	10690
	8.10 H_2O	770
	Pressure = 375 PSIA	
	Temperature = 120 ^oF	
9.	Sulfur	5800

BFW/Stm Stream	Mtl	PSIA	^oF	lb/Hr
11.	BFW	1600	274	54400
12.	BFW	1600	591	237800
13.	STM	1570	603	283900
14.	STM	225	392	15100
15.	STM	65	298	47700

Heat Flow	Avail Temp ^oF	Avail MBtu/Hr
Resaturator (Q162)	300	45400
BFW (Q163)	185	61400

Power Balance	MW
Generation	235.4
Gas Turbine	137.9
Stm Turbine	97.5
Use	23.5
Gasification	1.9
Oxygen	18.0
Station Power	3.6
Net to Grid	211.9

Heat Rate	Overall Efficiency
8278 Btu/kWHr (HHV)	41.22% (HHV)
7946 Btu/kWHr (LHV)	42.95% (LHV)

TABLE 2
MATERIAL AND ENERGY BALANCE
ONE MODULE - ~200 MW GAS IGCC

Coal Source: Donkin (Tunnel) Mine / Harbour Seam

MATERIAL

Process Stream		lb/Hr
1.	Coal	131900
	1.1 Dry Solids	123100
	Moisture	8800
2.	Limestone	None
3.	95% Oxygen	107900
	3.1 O_2	102700
	3.2 N_2 + A	5200
4.	Nitrogen	2000
5.	Water	75300
6.	Slag	13700
	6.1 Dry Solids	10200
	6.2 Moisture	3500
7.	Water Purge	47700
8.	Sour Syngas	259483
	8.1 H_2S	4720
	8.2 H_2	9620
	8.3 NH_3	3
	8.4 HCN	10
	8.5 CO	180250
	8.6 COS	360
	8.7 CH_4	1390
	8.8 CO_2	53470
	8.9 N_2 + A	8890
	8.10 H_2O	770
	Pressure = 375 PSIA	
	Temperature = 109 °F	
9.	Sulfur	4500

BFW/Stm Stream	Mtl	PSIA	°F	lb/Hr
11.	BFW	1600	274	53800
12.	BFW	1600	591	195400
13.	STM	1570	603	242300
14.	STM	225	392	13100
15.	STM	65	298	51200

Heat Flow	Avail Temp °F	Avail MBtu/Hr
Resaturator (Q162)	300	40300
BFW (Q163)	169	48500

Power Balance	MW
Generation	229.1
Gas Turbine	136.5
Stm Turbine	92.6
Use	19.4
Gasification	1.5
Oxygen	14.4
Station Power	3.5
Net to Grid	209.8

Heat Rate
7987 Btu/kWHr (HHV)
7708 Btu/kWHr (LHV)

Overall Efficiency
42.73% (HHV)
44.28% (LHV)

TABLE 3
MATERIAL AND ENERGY BALANCE
ONE MODULE - ~200 MW GAS IGCC

Coal Source: Phalen - Lingan Mine / Phalen - Harbour Seam

MATERIAL

Process Stream		lb/Hr
1.	Coal	123000
	1.1 Dry Solids	112500
	Moisture	10500
2.	Limestone	None
3.	95% Oxygen	101300
	3.1 O_2	96300
	3.2 N_2 + A	5000
4.	Nitrogen	2000
5.	Water	75200
6.	Slag	8200
	6.1 Dry Solids	6100
	6.2 Moisture	2100
7.	Water Purge	47900
8.	Sour Syngas	249143
	8.1 H_2S	2390
	8.2 H_2	9680
	8.3 NH_3	3
	8.4 HCN	20
	8.5 CO	178010
	8.6 COS	180
	8.7 CH_4	1630
	8.8 CO_2	47800
	8.9 N_2 + A	8670
	8.10 H_2O	760
	Pressure = 375 PSIA	
	Temperature = 110 °F	
9.	Sulfur	2300

BFW/Stm Stream	Mtl	PSIA	°F	lb/Hr
11.	BFW	1600	274	53200
12.	BFW	1600	591	187100
13.	STM	1570	603	235000
14.	STM	225	392	12700
15.	STM	65	298	44100

Heat Flow	Avail Temp °F	Avail MBtu/Hr
Resaturator (Q162)	298	37000
BFW (Q163)	156	41200

Power Balance	MW
Generation	227.8
Gas Turbine	136.1
Stm Turbine	91.7
Use	18.0
Gasification	1.0
Oxygen	13.5
Station Power	3.5
Net to Grid	209.8

Heat Rate
7836 Btu/kWHr (HHV)
7572 Btu/kWHr (LHV)

Overall Efficiency
43.55% (HHV)
45.07% (LHV)

FROM: M.J.PFEIFFER (MFTN/106)
F.A.M. SCHRIJVERS (MFTN/104)

28 APRIL 1989.

TO : MR. DONALD ANDERSON

RESEARCH ENGINEER

ATLANTIC COAL INSTITUTE.

REF: FAX 30/3/89.

SUBJECT: INTEGRATED GASIFICATION COMBINED CYCLE.

WE ARE PLEASED TO PROVIDE YOU WITH THE FOLLOWING NON-CONFIDENTIAL INFORMATION IN RESPONSE TO YOUR REQUEST FOR MASS AND HEAT BALANCE DATA FOR THE SHELL COAL GASIFICATION PROCESS (SCGP).

AAA) BASIS OF THE STUDY.

THE CALCULATIONS ARE BASED ON THE FOLLOWING COAL PROPERTIES, WHICH HAVE BEEN DERIVED FROM THE INFORMATION RECEIVED FROM YOU:

PROXIMATE ANALYSIS (AR) % WT:

FIXED CARBON	48.66 % WT
VOLATILE MATTER	33.02 "
MOISTURE	6.71 "
ASH	11.61 "
TOTAL	100.00 % WT

ULTIMATE ANALYSIS (AR) % WT:

CARBON	66.78 % WT
HYDROGEN	5.69 "
OXYGEN	10.49 "
SULPHUR	4.07 "
NITROGEN	1.36 "
ASH	11.61 "
TOTAL	100.00 % WT

NET HEAT OF COMBUSTION (AR) : 27039 MJ/TON.

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BBB) MASS BALANCE.

FOR CLARITY REASONS WE SPLIT UP THE MASS BALANCE OVER THE FOLLOWING THREE SECTIONS: COAL MILLING AND DRYING, GASIFICATION AND DRY PARTICULATE REMOVAL, AND WET PARTICULATE REMOVAL AND SOUR WATER TREATING.

THE CALCULATIONS ARE DONE FOR A COAL THROUGHPUT OF 1000 T/D PRINCE HUB COAL. FOR DIFFERENT THROUGHPUTS OF PRINCE HUB COAL YOU MAY SCALE LINEARLY FROM THE DATA GIVEN.

1) COAL MILLING AND DRYING:

	T/D	P(BARA)	T(DEGR.C)
IN: - COAL INTAKE (AS RECEIVED)	1,000	1.0	AMBIENT
- NATURAL GAS FOR COAL DRYING	5	5.5	"
- AIR	185	1.3	"
TOTAL IN	1,190		
OUT: - BLEED STREAM FLUE GASES	238	1.1	100
DRIED COAL TO GASIFIER	952		
TOTAL OUT	1,190		

2) GASIFICATION AND DRY PARTICULATE REMOVAL.

	T/D	P(BARA)	T(DEGR.C)
IN: - COAL FROM DRIER	952		
- TRANSPORT N2	60	28.0	100
- OXYGEN TO GASIFIER	851	28.0	250
- MP STEAM TO GASIFIER	89	35.0	242
- WATER TO SLAG COOLING	20	1.0	35
TOTAL IN	1,972		
OUT: SYNTHESIS GAS TO WET SOLIDS REMOVAL	1,832		
SLAG AND FLYSLAG (*)	140	1.0	60
TOTAL OUT	1,972		

(*) INCLUDING WATER AND UNCONVERTED CARBON ON FLY SLAG.

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CARBON CONVERSION IN SCGP AMOUNTS TO 99.5 % IN THIS CASE. THE CARBON CONVERSION INCREASES WHEN FLY SLAG RECYCLE IS APLLIED. IN THAT CASE THE CARBON CONVERSION WILL BE APPROXIMATELY 99.9 %.

DURING NON FLY SLAG RECYCLE OPERATION APPROXIMATELY 70.0 % OF THE ASH IN THE COAL COMES OUT AS (BOTTOM) SLAG AND 30.0 % AS FLY SLAG. IN CASE OF FLY SLAG RECYCLE APPROXIMATELY 95.0 % OF THE ASH LEAVES THE GASIFIER AS SLAG AND ONLY 5.0 % AS FLY SLAG.

FLY SLAG AND SLAG ARE NON-LEACHABLE BY-PRODUCTS SUITABLE FOR USE IN COMMERCIAL CONSTRUCTION MATERIAL.

FOR PROPER OPERATION OF THE LOCK HOPPER SYSTEM AN ADDITIONAL HIGH PRESSURE NITROGEN (50.0 - 60.0 BARA) CONSUMPTION OF 185 T/D IS REQUIRED.

3) WET PARTICULATE REMOVAL AND SOUR WATER TREATING

	T/D	P(BARA)	T(DEGR.C)
IN: SYNTHESIS GAS TO WET SOLIDS REMOVAL	1,832		
- LP STEAM	182	6.0	159
TOTAL IN	2,014		
OUT: - SYNTHESIS GAS	1,792	21.0	40
- SOUR WATER STRIPPER OVERHEAD	2	2.0	40
- PROCESS WATER	220	3.0	40
TOTAL OUT	2,014		

CCC) HEAT BALANCE

		GJ/D
IN:	- COAL INTAKE	27,039
	- NATURAL GAS FOR COAL DRYING	217
	- MP STEAM	250
	- LP STEAM	502
	- INTERNAL POWER CONSUMPTION	119
	TOTAL IN	28,127
OUT:	- SYNTHESIS GAS	22,319
	- HIGH LEVEL HEAT AVAILABLE	4,283
	- LOW LEVEL HEAT AVAILABLE	1,003
	- UNCONVERTED CARBON	135
	- HEAT LOSSES	387
	TOTAL OUT	28,127

THE COLD GAS EFFICIENCY (DEFINED AS LOWER HEATING VALUE OF THE UNTREATED SYNTHESIS GAS DIVIDED BY THE LHV OF THE COAL TIMES 100.0 %) IS 82.5%. THE COLD GAS EFFICIENCY OF THE TREATED SYNGAS TO GAS TURBINE (AFTER SULPHUR REMOVAL) IS 80.4 %.

NOTE 1: THE ACTUAL STEAM QUALITY DEPENDS ON POWER BLOCK DESIGN AND DEGREE OF INTEGRATION WITH SCGP. A STEAM PRESSURES OF 110 BARA AND TEMPERATURE OF 540 DEGR.C IS TYPICAL FOR THE SYNGAS COOLER. A TYPICAL STEAM PRESSURE IN THE GASIFIER MEMBRAME WALL IS 40.0 BARA, PRESSURES UP TO 70 BARA ARE POSSIBLE IF DESIRED. IN THE MEMBRAME WALL 543 GJ/D OF THE TOTAL HIGH LEVEL HEAT IS RECOVERED BY EVAPORATING SURFACES.

THE REMAINING HIGH LEVEL HEAT IS AVAILABLE AT A TEMPERATURE LEVEL OF 900 - 230 DEGR.C FOR HIGH QUALITY STEAM GENERATION. IN THE TEMPERATURE INTERVAL FROM 900 - 780 DEG C THE REMAINING HEAT SHOULD ALSO BE REMOVED BY EVAPORATING SURFACES.

NOTE 2: LOW LEVEL HEAT IS AVAILABLE AT THE FOLLOWING TEMPERATURE LEVELS:

125 - 49 DEGR.C	563	GJ/D
95 60 DEGR.C	125	"
95 - 40 DEGR.C	315	"

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DDD) SYNTHESIS GAS COMPOSITION/PROPERTIES

THE SYNTHESIS GAS PRODUCED FROM THE PRINCE HUB COAL USING 95 VOL % OXYGEN HAS THE FOLLOWING COMPOSITION BEFORE (I) AND AFTER (II) TREATMENT FOR SOUR GAS REMOVAL:

	I	II
H2	30.6	31.3
CO	61.6	62.8
CO2	1.5	1.0
H2S	1.3	(*)
COS	0.2	(*)
CH4	100 ppm	100 ppm
N2	3.4	3.4
Ar	1.1	1.1
H2O	0.3	0.4
TOTAL	100.0	100.0

NOTE 3:(*) SULPHUR IN THE FORM OF COS AND H2S TOGETHER : 300 PPM.

THE MASS FLOW OF TREATED SYNGAS AFTER SOUR GAS REMOVAL: 1730 TON/DAY

THE LHV OF TREATED SYNGAS : 12.57 MJ/KG.

PRESSURE : 20.0 BARA.

TEMPERATURE OF TREATED SYNGAS : 40.0 DEG C.

AN OVERALL SULPHUR REMOVAL EFFICIENCY OF 98 % WAS ASSUMED, WHICH RESULTS IN A SALEABLE SULPHUR PRODUCTION OF ABOUT 39.9 KG SULPUR PER TON COAL (AR).

BASED ON THE CALCULATIONS AND THE COAL AND ASH ANALYSES AVAILABLE WE CONSIDER THE SHELL COAL GASIFICATION PROCESS SUITABLE FOR HANDLING THE PRINCE HUB COAL. THE OTHER COALS (DONKIN HARBOUR AND PHALEN/LINGAN) SEEM ALSO TO BE SUITABLE FOR OUR PROCESS BASED ON THE COAL AND ASH ANALYSES AS GIVEN IN YOUR FAX.

WE HOPE THE ABOVE SUITS YOUR NEEDS FOR THE PRESENT. WE ARE LOOKING FORWARD TO RECEIVE THE RESULTS OF YOUR STUDY.

THANK YOU FOR YOUR INTEREST IN OUR COAL GASIFICATION TECHNOLOGY.

REGARDS

F.A.M. SCHRIJVERS / M.J.PFEIFFER.